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Treatability Manual. Volume IV
Cost Estimating

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The objective of the treatability program are: To provide readily accessible data and information on treatability of industrial and municipal waste streams for use by NPDES permit writers, enforcement personnel, and by industrial or municipal permit holders; To provide a basis for research planning by identifying gaps in knowledge of the treatability of certain pollutants and wastestreams; To set up a system allowing rapid response to program office requirements for generation of treatability data.

KEYWORDS: *Waste water, *Water pollution control.

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TREATABILITY MANUAL

VOLUME IV. Cost Estimating

OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C. 20460

July 1980

REPRODUCED BY
NATIONAL TECHNICAL
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U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA 22161

PREFACE

In January, 1979, EPA's Office of Enforcement and Office of Water and Waste Management requested help from the Office of Research and Development in compiling wastwater treatment performance data into a "Treatability Manual". This Manual was to be used in developing NPDES permit limitations for facilities which, at the time of permit issuance, were not fully covered by promulgated, industry-specific effluent guidelines authorized under Sections 301, 304, 306, 307, and 501 of the CWA.

A planning group was set up to manage the treatability program under the chairmanship of William Cawley, Deputy Director, Industrial Environmental Research Laboratory - Cincinnati. The group includes participants from: 1) the Industrial Environmental Research Laboratory - Cincinnati, 2) Effluent Guidelines Division, Office of Water and Waste Management; 3) Permits Division, Office of Enforcement; 4) Municipal Environmental Research Laboratory - Cincinnati; 5) R. S. Kerr, Environmental Research Laboratory - Ada; 6) Industrial Environmental Research Laboratory - Research Triangle Park; 7) Monsanto Research Corporation; and 8) Aerospace Corporation.

The objectives of the treatability program are:

- To provide readily accessible data and information on treatability of industrial and municipal waste streams for use by NPDES permit writers, enforcement personnel, and by industrial or municipal permit holders;
- To provide a basis for research planning by identifying gaps in knowledge of the treatability of certain pollutants and wastestreams;
- To set up a system allowing rapid response to program office requirements for generation of treatability data.

The primary output from this program is a five-volume Treatability Manual. The individual volumes are named as follows:

Volume I - Treatability Data
Volume II - Industrial Descriptions
Volume III - Technologies
Volume IV - Cost Estimating
Volume V - Summary

Date: 6/23/80

ACKNOWLEDGEMENT

The sheer size and comprehensiveness of this document should make it obvious that this had to be the effort of a large number of people. It is the collection of contributions from throughout the Environmental Protection Agency, particularly from the Office of Enforcement, Office of Water and Hazardous Materials and the Office of Research and Development. Equally important to its success were the efforts of the employees of the Aerospace Corporation, MATHTECH, INC., and the Monsanto Research Corporation who participated in this operation.

No list of the names of everyone who took part in the effort would in any way adequately acknowledge the effort which those involved in preparing this Manual made toward its development. Equally difficult would be an attempt to name the people who have made the most significant contributions both because there have been too many and because it would be impossible to adequately define the term "significant." This document exists because of major contributions by the contractor's staff and by members of the following:

Effluent Guidelines Division
    Office of Water and Waste Management

Permits Division
    Office of Water Enforcement

National Enforcement Investigation Center
    Office of Enforcement

Center for Environmental Research Information

Municipal Environmental Research Laboratory

Robert S. Kerr Environmental Research Laboratory

Industrial Environmental Research Laboratory
    Research Triangle Park, NC

Industrial Environmental Research Laboratory
    Cincinnati, OH
    Office of Research and Development

The purpose of this acknowledgement is to express my thanks as Committee Chairman and the thanks of the Agency to the Committee Members and others who contributed to the success of this effort.

William A. Cawley, Deputy Director, IERL-Ci
Chairman, Treatability Coordination Committee

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GLOSSARY

AAP: Army Ammunitions Plant.

AN: Ammonium Nitrate.

ANFO: Ammonium Nitrate/Fuel Oil.

BATEA: Best Available Technology Economically Achievable.

BAT: Best Applicable Technology.

BDL: Below Detection Limit.

BEJ: Best Engineering Judgement.

BOD: Biochemical Oxygen Demand.

clarification: Process by which a suspension is clarified to give a "clear" supernatant.

cryolite: A mineral consisting of sodium-aluminum fluoride.

CWA: Clean Water Act.

cyanidation process: Gold and/or silver are extracted from finely crushed ores, concentrates, tailings, and low-grade mine-run rock in dilute, weakly alkaline solutions of potassium or sodium cyanide.

comminutor: Mechanical devices that cut up material normally removed in the screening process.

effluent: A waste product discharged from a process.

EGD: Effluent Guidelines Division.

elutriation: The process of washing and separating suspended particles by decantation.

extraction: The process of separating the active constituents of drugs by suitable methods.

fermentation: A chemical change of organic matter brought about by the action of an enzyme or ferment.

flocculation: The coalescence of a finely-divided precipitate.

fumigant: A gaseous or readily volatilizable chemical used as a disinfectant or pesticide.

GAC: Granular Activated Carbon.

gravity concentration: A process which uses the differences in density to separate ore minerals from gangue.

gravity separation/settling: A process which removes suspended solids by natural gravitational forces.

grit removal: Preliminary treatment that removes large objects, in order to prevent damage to subsequent treatment and process equipment.

influent: A process stream entering the treatment system.

intake: Water, such as tap or well water, that is used as makeup water in the process.

lagoon: A shallow artificial pond for the natural oxidation of sewage or ultimate drying of the sludge.

LAP: Loading Assembly and Packing operations.

MGD: Million Gallons per Day.

MHF: Multiple Hearth Furnace.

NA: Not Analyzed.

ND: Not Detected.

neutralization: The process of adjusting either an acidic or a basic wastestream to a pH near seven.

NPDES: National Pollutant Discharge Elimination System.

NRDC: National Resources Defense Council.

NSPS: New Source Performance Standards.

photolysis: Chemical decomposition or dissociation by the action of radiant energy.

PCB: PolyChlorinated Biphenyl.

POTW: Publicly Owned Treatment Works.

PSES: Pretreatment Standards for Existing Sources.

purged: Removed by a process of cleaning; take off or out.

screening process: A process used to remove coarse and/or gross solids from untreated wastewater before subsequent treatment.

SIC: Standard Industrial Classification.

SS: Suspended Solids.

SRT: Solids Retention Time.

starved air combustion: Used for the volumetric and organic reduction of sludge solids.

terpene: Any of a class of isomeric hydrocarbons.

thermal drying: Process in which the moisture in sludge is reduced by evaporation using hot air, without the solids being combusted.

TKN: Total Kjeldahl Nitrogen.

TOC: Total Organic Carbon.

trickling filter: Process in which wastes are sprayed through the air to absorb oxygen and allowed to trickle through a bed of rock or synthetic media coated with a slime of microbial growth to removed dissolved and collodial biodegradable organics.

TSS: Total Suspended Solids.

vacuum filtration: Process employed to dewater sludges so that a is produced having the physical handling characteristics and contents required for processing.

VSS: Volatile Suspended Solids.

WQC: Water Quality Criterion.

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IV.1 INTRODUCTION

This manual presents total capital investment and annual operating cost curves for wastewater treatment technologies. The cost curves, which generally show costs in millions of dollars versus wastewater flowrate, are presented for 78 technologies grouped into the following six classifications: wastewater conditioning, primary wastewater treatment, secondary wastewater treatment, tertiary wastewater treatment, sludge treatment, and disposal.

Each technology is described in terms of a general overview, discussion of common modifications and typical equipment, and a flow diagram. Design criteria are briefly discussed and cost assumptions for each technology are presented. As discussed in Appendix A-Economic Assumptions, all costs are indexed to September 1979 (corresponding to an Engineering News Record index of 3119) unless noted.

Cost data presented in this manual are generalized rather than site specific, and are intended primarily for comparative analysis. The user of this manual is cautioned against treating any cost estimates based on the data contained herein as absolute. These estimates are valid only for comparison purposes, and even these comparisons must be performed with caution because of the possible differences in the reliability of the performance and cost information from various sources. Further, there may be considerable difference between cost estimates for comparative purposes and the actual total capital investment for the facilities.

Cost data presented in this manual have been derived from EPA publications, open literature, construction grant files, and manufacturers' information and have been indexed to 1979. These data exhibit a level of accuracy dependent upon the degree of usage of the process. Higher reliability costs generally occur where a large amount of historical data exists, such as clarifiers and activated sludge. Those processes with few examples beyond the demonstration stage must be considered as individual cases with the potential for wider variation in costs when applied in a generalized fashion. It can generally be said that the data on a specific control alternative are sufficient only to establish a relationship as expressed by a table of data or a single curve or family of curves (± approximately 30%).

A large number of the cost curves developed for this manual have been obtained or derived from the Areawide Assessment Procedures Manual (EPA-600/9-76-014) and Innovative and Alternative Technology Assessment Manual (EPA-430/9-78-009). This information is the result of cost estimates developed over a period of years by several contractors from detailed conceptual designs, process and equipment layouts in accordance with standard estimating techniques, plus verification using normalized "as built" costs where available. Processes having limited "as built" cost data bases contain single or multiple case history costs as available. In some cases, costs have been tabulated rather than displayed graphically.

Land costs, except in land application and landfill, are not included in the cost curves. Land costs are highly dependent on local conditions and should be developed on a case-by-case basis. It should be noted, however, that land costs (except for land application and land disposal of sludge) may not be necessary for comparison purposes.

Cost components included in the estimation of the total capital investment are shown below. Cost components included in the estimation of total annual operating costs are shown on the following page. A more detailed discussion of economic assumptions is presented in Appendix A.

BREAKDOWN OF TOTAL CAPITAL INVESTMENT

FIXED CAPITAL INVESTMENT

Direct cost components:

- Purchased equipment and installation
- Instrumentation and controls
- Piping
- Electrical equipment and materials
- Buildings
- Yard improvements
- Service facilities

Indirect cost components:

- Engineering and supervision
- Construction expenses
- Contractors fee
- Contingency

WORKING CAPITAL

BREAKDOWN OF ANNUAL OPERATING COST

TOTAL ANNUAL OPERATING COST:

Total direct operating cost:

- Labor
- Materials
- Chemicals
- Power
- Fuel

Total indirect operating cost:

- Plant overhead
- Taxes and insurance
- General and administrative expenses
- Depreciation
- Interest on working capital

IV.2 WASTEWATER CONDITIONING

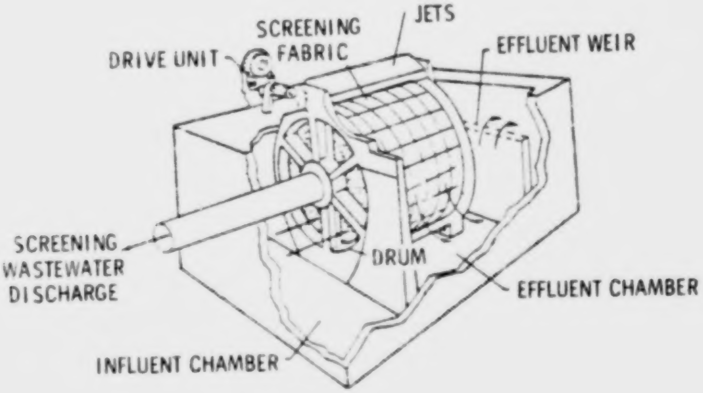
IV.2.1 SCREENING

IV.2.1.1 Horizontal Rotary Shaft Screen [1]

Description. A horizontal rotary-shaft screen is an intermittently or continuously rotating drum covered with a plastic or stainless-steel screen of uniform-sized openings, installed and partially submerged in a chamber. The chamber is designed to permit entry of wastewater to the interior of the drum and collection of filtered (or screened) wastewater from the exterior side of the drum. With each revolution, the solids are flushed by sprays from the exposed screen surface into a collecting trough. Different types of screen are used. Coarse screens have openings of 1/4 inch or more; fine screens have openings less than 1/4 inch; screens with openings of 20 to 70 μm are called microscreens or microstrainers. Drum diameters are usually 3 to 5 feet; lengths are 4 to 12 feet. The most common modifications for horizontal rotary shaft screen are the following:

Tile chambers; reinforced concrete chamber; steel chamber
Variable-speed drive for drum
Addition of backwash storage and pumping facilities
Addition of ultraviolet light slime-growth control equipment
Addition of chlorinating equipment

Equipment normally associated with the horizontal rotary-shaft screen process includes screens and mechanical components. A flow diagram of a horizontal rotary-shaft screen is presented below.



Typical design criteria. Typical design criteria of horizontal rotary-shaft screen are shown below.

Design criterion	Value
Screen submergence	70 - 80%
Loading rate	2 - 10 gpm/ft^2 of submerged area, depending on pretreatment and mesh size
Screen openings	150 μm - 0.4 in. for pretreatment; 20 - 70 μm for tertiary treatment
Drum rotation	0 - 7 revolutions/min
Screen materials	Stainless steel or plastic cloth
Washwater quantity	2 - 5% of flow being treated
Performance of fine screen device	Varies considerably depending on influent solids type, concentration and loading patterns, mesh size, hydraulic head, and degree of biological conditioning of solids

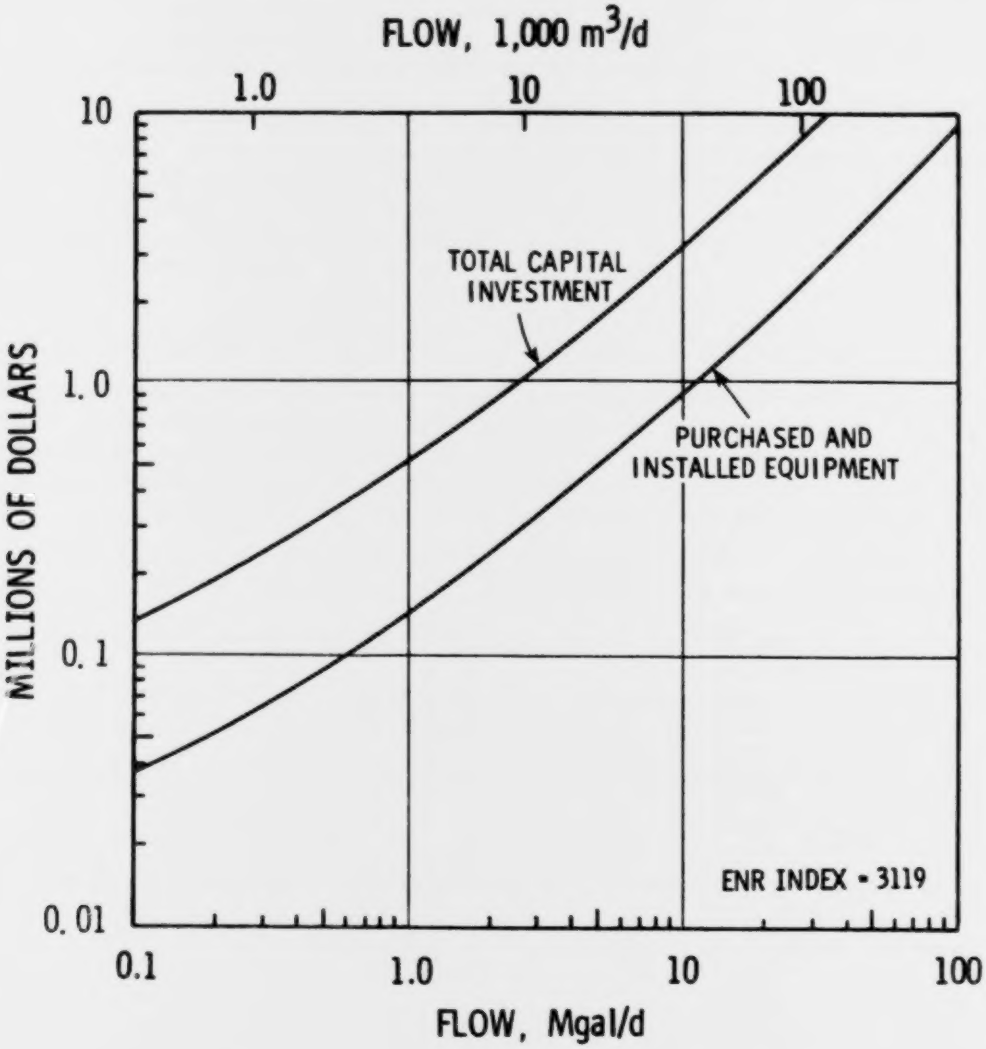
Costs. Purchased equipment and installation cost for estimation of total capital investment includes tanks, drums, screens, backwash equipment, drive motors, and building. Instrumentation for automatic operation is also included. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Hydraulic load	2.5 gpm/ft^2 at average flow
Screen mesh	25 μ
Peripheral drum speed	15 ft/min at 3-in. head loss
Backwash	3% of throughput at 35 psi

References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



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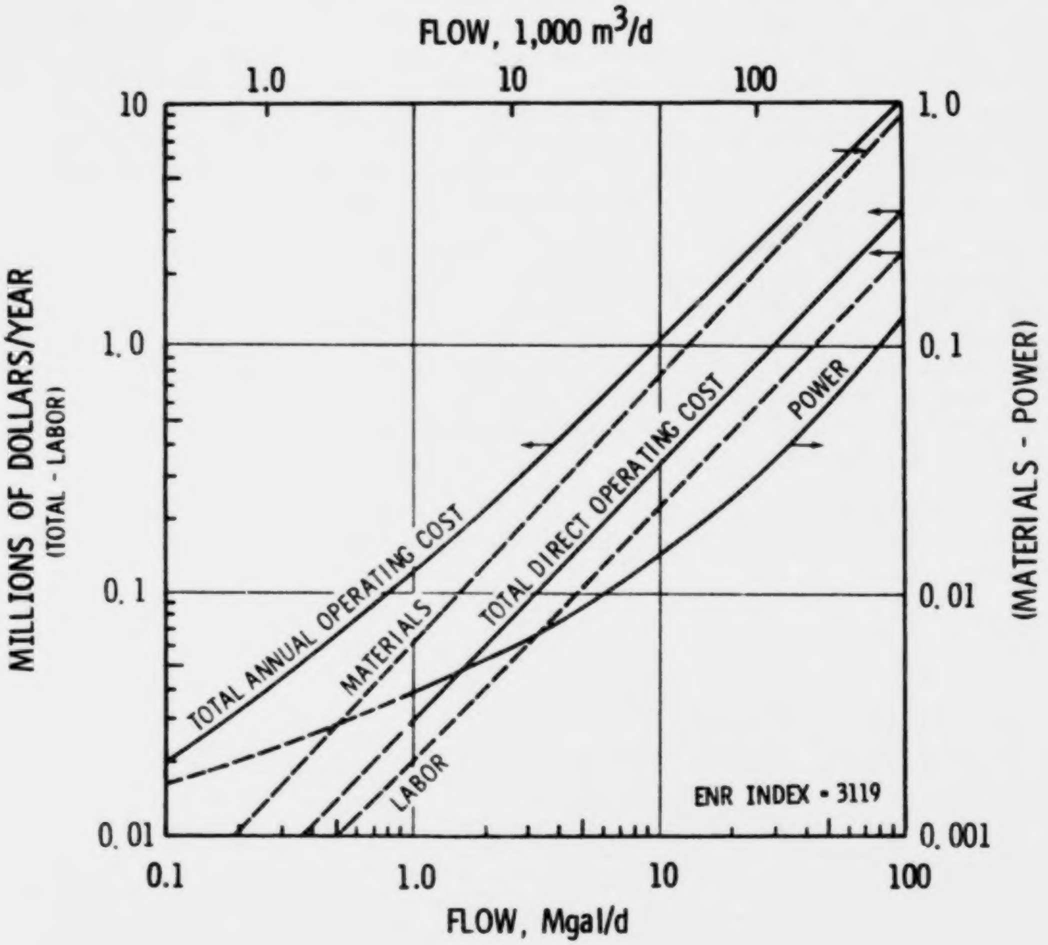
ANNUAL OPERATING COST

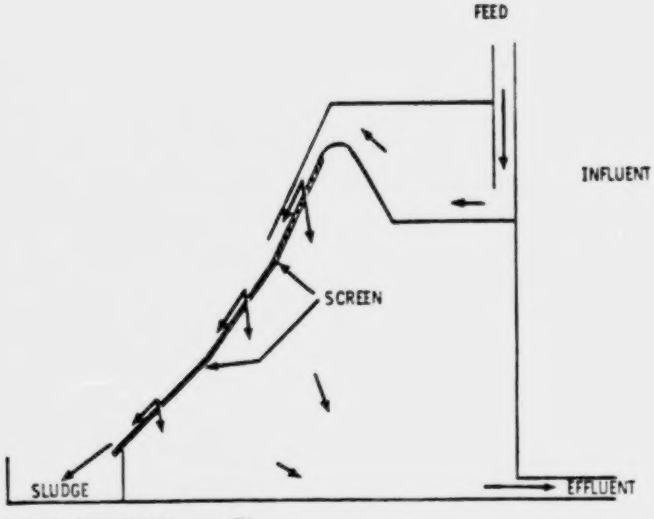


IV.2.1.2 Wedge-wire Screen [1]

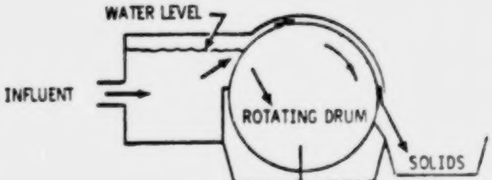
Description. A wedge-wire screen is a device onto which wastewater is directed across an inclined stationary screen or a drum screen of uniform-sized openings. Solids are trapped on the screen surface while the wastewater flows through the openings. The solids are moved either by gravity (stationary) or by mechanical means (rotating drum) to a collecting area for discharge. Stationary screens introduce the wastewater as a thin film flowing downward with a minimum of turbulence across the wedge-wire screens, which are generally in three sections of progressively flatter slope. The drum screen employs the same type of wedge wire wound around its periphery. Wastewater is introduced as a thin film near the top of the drum and flows through the hollow drum and out the bottom. The solids retained by the peripheral screen follow the drum rotation until removed by a doctor blade located at about 120° from the introduction point.

Wedge-wire spacing can be varied to best suit the application. For municipal wastewater applications, spacings are generally between 0.01 and 0.06 inches (0.25 to 1.5 mm). Inclined screens can be housed in stainless steel or fiberglass; wedge-wires may be curved or straight; the screen face may be a single multi-angle unit, three separate multi-angle pieces, or a single curved unit. Rotary screens can have a single rotation speed drive or a variable speed drive.

Equipment normally associated with this technology includes screen systems. A flow diagram of a wedge-wire screen system is shown below.



(SEE SECTION IV.7)



(SEE SECTION IV.7)

Typical design criteria. Typical design criteria of a wedge-wire screen based on screening of raw wastewater at 0.05 - 36 Mgal/d are shown below.

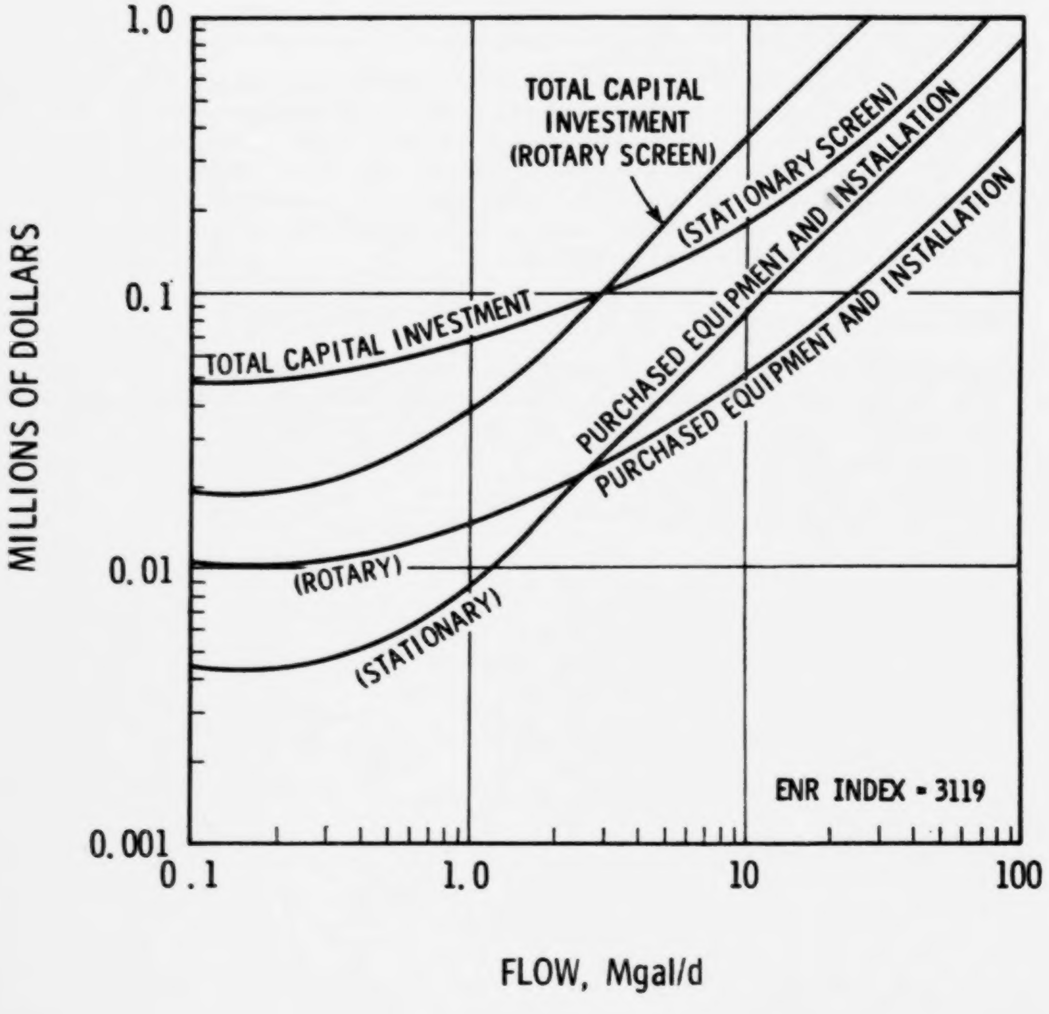
Design criterion	Value Stationary screen	 Rotary drum screen
Screen opening	0.01 - 0.06 in.	0.01 - 0.06 in.
Head required	4 - 7 ft	2.5 - 4.5 ft
Space required	10 - 750 ft^2	10 - 100 ft^2
Motor size	-	0.5 - 3 hp

Costs. Purchased equipment and installation cost for estimation of total capital investment includes wedge-wire stainless-steel screen with 0.06 in. opening, electrical provisions, and equipment that provides suitable weirs for flow control. Flumes, piping for effluent or sludge, and pumping equipment are not included. Annual operating costs are based on the pumping head of 4.5 ft for a stationary screen.

References.

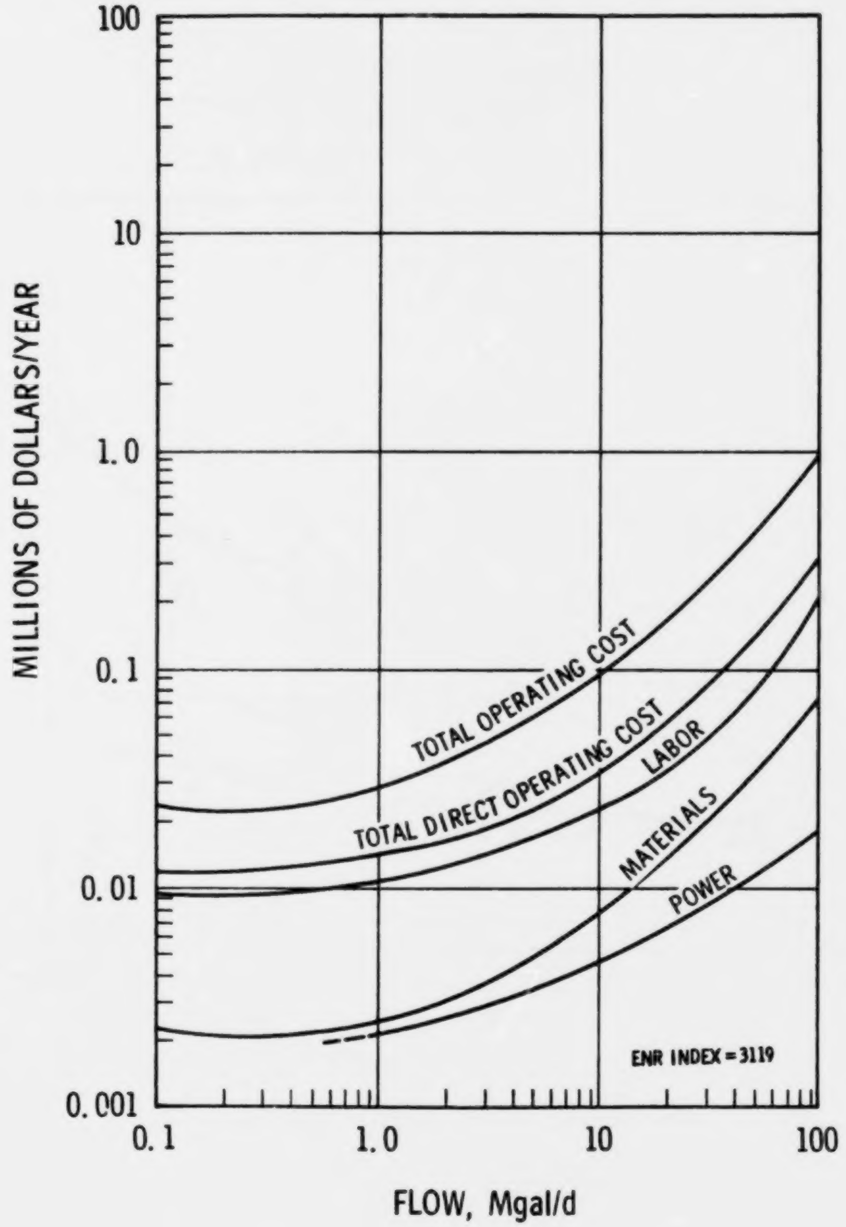
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT

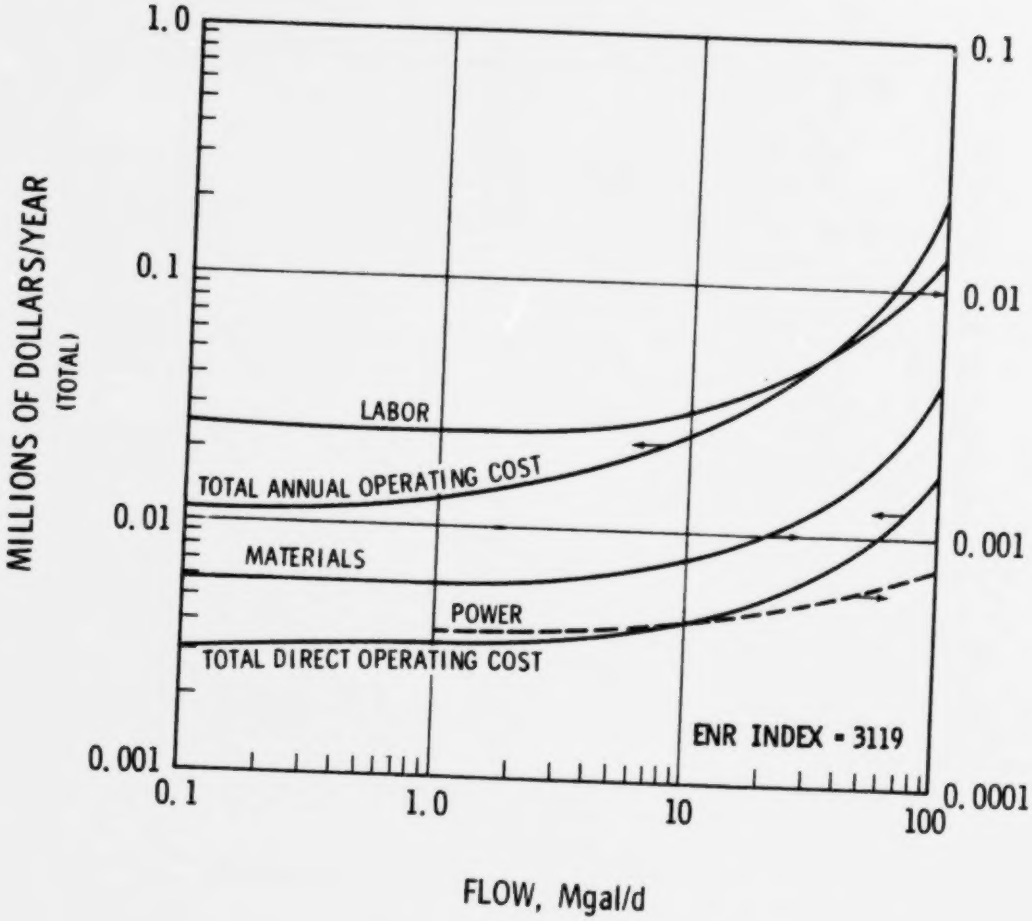


ANNUAL OPERATING COST

(STATIONARY SCREEN)



ANNUAL OPERATING COST
(ROTARY SCREEN)



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IV.2.2 GRIT REMOVAL (PRELIMINARY TREATMENT) [1]

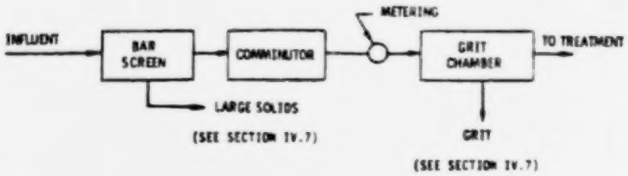
IV.2.2.1 Description

Preliminary treatment usually consists of two separate and distinct unit operations: bar screening and grit removal. There are two types of bar screens (or racks). The most commonly used, and oldest technology, consists of hand-cleaned bar racks, which are generally used in smaller treatment plants. The second type of bar screen is the type that is mechanically cleaned, which is commonly used in larger facilities.

Grit is most commonly removed in chambers that are capable of settling out high density solid materials, such as sand, gravel, and cinders. There are two types of grit chambers: horizontal flow, and aerated; in both types the settleables collect at the bottom of the unit. Horizontal units are designed to maintain a relatively constant velocity by use of proportional weirs or flumes in order to prevent settling of organic solids, while simultaneously obtaining relatively complete removal of inorganic particles (grit). Aerated grit chambers produces spiral action whereby the heavier particles remain at the bottom of the tank to be removed, while organic particles are maintained in suspension by rising air bubbles. One main advantage of aerated units is that the amount of air can be regulated to control the grit/organic solids separation, and less offensive odors are generated. The aeration process also facilitates cleaning of the grit. Grit removed from horizontal flow units usually needs additional cleaning steps prior to disposal.

Some plants also use comminutors, which are mechanical devices that cut up the material normally removed in the screening process. Therefore, these solids remain in the wastewater to be removed in downstream unit operations, rather than being removed immediately from the wastewater. In recent years, the use of static or rotating wedge-wire screens has increased to remove large organic particulates just prior to degritting. These units have been found to be superior to comminutors in that they remove the material immediately from the waste instead of creating additional loads downstream. Other grit chamber designs are available including swirl concentrators and square tanks.

Equipment normally associated with preliminary treatment includes screens, grinders, comminutors, sedimentation equipment, and wedge-wire screens. A flow diagram for preliminary treatment including grit removal is shown below.



IV.2.2.2 Typical Design Criteria

Typical design criteria for screening and grit chambers are shown below.

Design criterion	Value
Bar screens	
Bar size	1/4 - 5/8 in. width by 1 - 3 in. depth
Spacing	3/4 - 3 in.
Slope from vertical	0 - 45°
Velocity	1.5 - 3 ft/s
Wedge-wire screens	See Section IV.2.1
Grit chambers	
Horizontal velocities	1/2 - 1.25 ft/s
Length	Units are sufficiently long to settle lightest and smallest (usually 0.2 mm) grit particles with an additional factor of safety (up to 50%)
Weir crest position	Generally set 4 - 12 in. above bottom

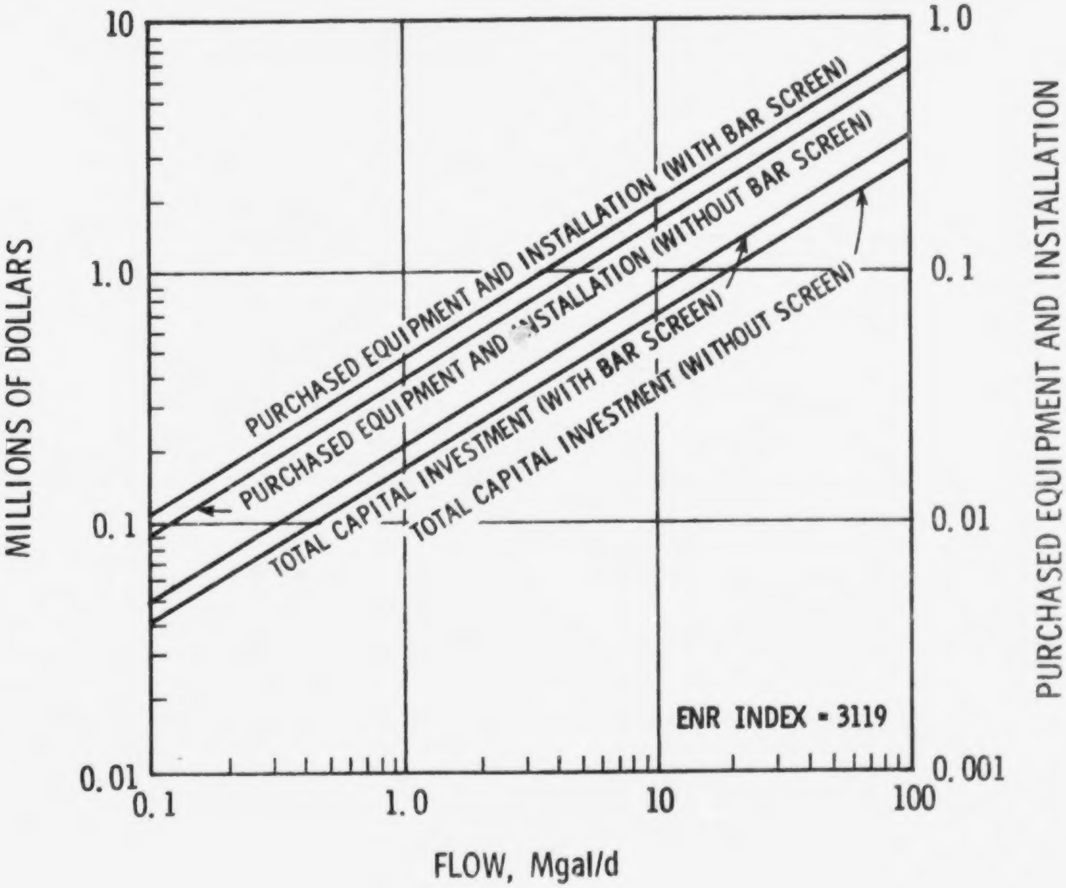
IV.2.2.3 Costs

Purchased equipment and installation cost includes flow channels and superstructures, bar screens (mechanical), horizontal grit chamber with mechanical grit handling equipment, Parshall flume, and flow-recording equipment. Total annual costs do not include cost for grit disposal. Screenings 1 to 3 ft^3 of suspended material/Mgal. Grit 2 to 5 ft^3 of grit/Mgal.

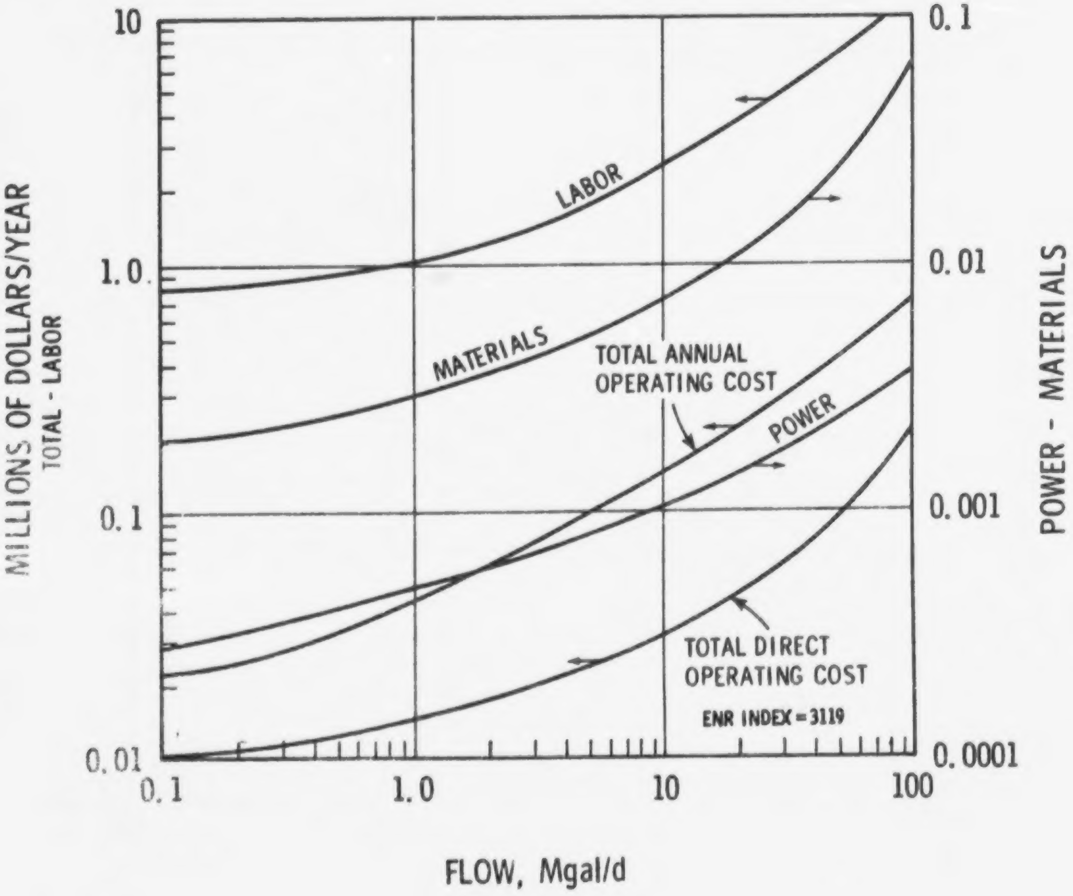
IV.2.2.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.2.3 FLOW EQUALIZATION [1]

IV.2.3.1 Description

Wastewater flows into treatment facilities are subjected to fluctuation in water quality and quantity. Most waste treatment processes are sensitive to such changes. An equalization basin serves to balance the extreme quality and quantity of these fluctuations to allow normal contact time in the treatment facility. This section only addresses equalization basins that are used to equalize flow; however, equalization of the quality of wastewater will also occur to a degree.

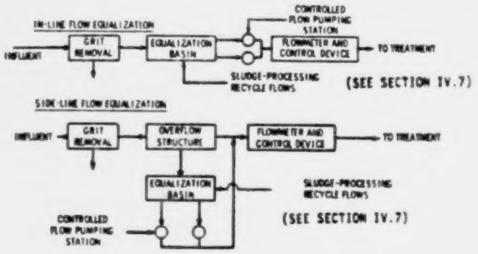
Equalization basins may be designed as either in-line or side-line units. In the in-line design, the basin recieves the wastewater directly from the collection system, and the discharge from the basin through the treatment plant is kept essentially at constant rate. In the side-line design, flows in excess of the average are diverted to the equalization basin and, when the plant flow falls below the average, wastewater from the basin is discharged to the plant to increase the flow to the average level. The basins are sufficiently sized to hold the peak flows and discharge at constant rate.

Pump stations may or may not be required to discharge into or out of the equalization basin, depending upon the available head. Where pumping is found necessary, the energy requirements are based on total flow for in-line basins and on excess flow for side-line basins.

Aeration of the wastewater in the equalization basin is normally required for mixing and maintaining aerobic conditions.

There are various methods of aeration, pumping, and flow control. Tanks or basins can be manufactured from steel or concrete, or they can be excavated and of the lined or unlined earthen variety.

Equipment normally associated with flow equalization includes lift pumps, air compressors, basin liners, flow controllers, and aerators. Flow diagrams for in-line and side-line flow equalization are shown below.


IN-LINE FLOW EQUALIZATION
INFLUENT
GRIT REMOVAL
EQUALIZATION BASIN
CONTROLLED FLOW PUMPING STATION
FLOWMETER AND CONTROL DEVICE
TO TREATMENT
SLUDGE-PROCESSING RECYCLE FLOWS
(SEE SECTION IV.7)
SIDE-LINE FLOW EQUALIZATION
INFLUENT
GRIT REMOVAL
OVERFLOW STRUCTURE
FLOWMETER AND CONTROL DEVICE
TO TREATMENT
EQUALIZATION BASIN
SLUDGE-PROCESSING RECYCLE FLOWS
(SEE SECTION IV.7)
CONTROLLED FLOW PUMPING STATION

IV.2.3.2 Typical Design Criteria

The design of an equalization basin is highly site specific and dependent upon the type and magnitude of the input flow variations and facility configurations. Typical design criteria and their dependencies are shown below for flow equalization.

Design criteria	Value
Pumping/flow control mode, aeration, mixing, and flushing methods	Dependent upon size and site conditions
Grit removal	Should be provided upstream of basin
Mechanical mixing[a]	20 - 40 hp/Mgal of storage
Aeration[a]	1.25 - 2 ft^3/min/1,000 gal of storage[b]

[a] Mechanical aerators will supply both mixing and aeration.

[b] Dependent upon IOD and BOD.

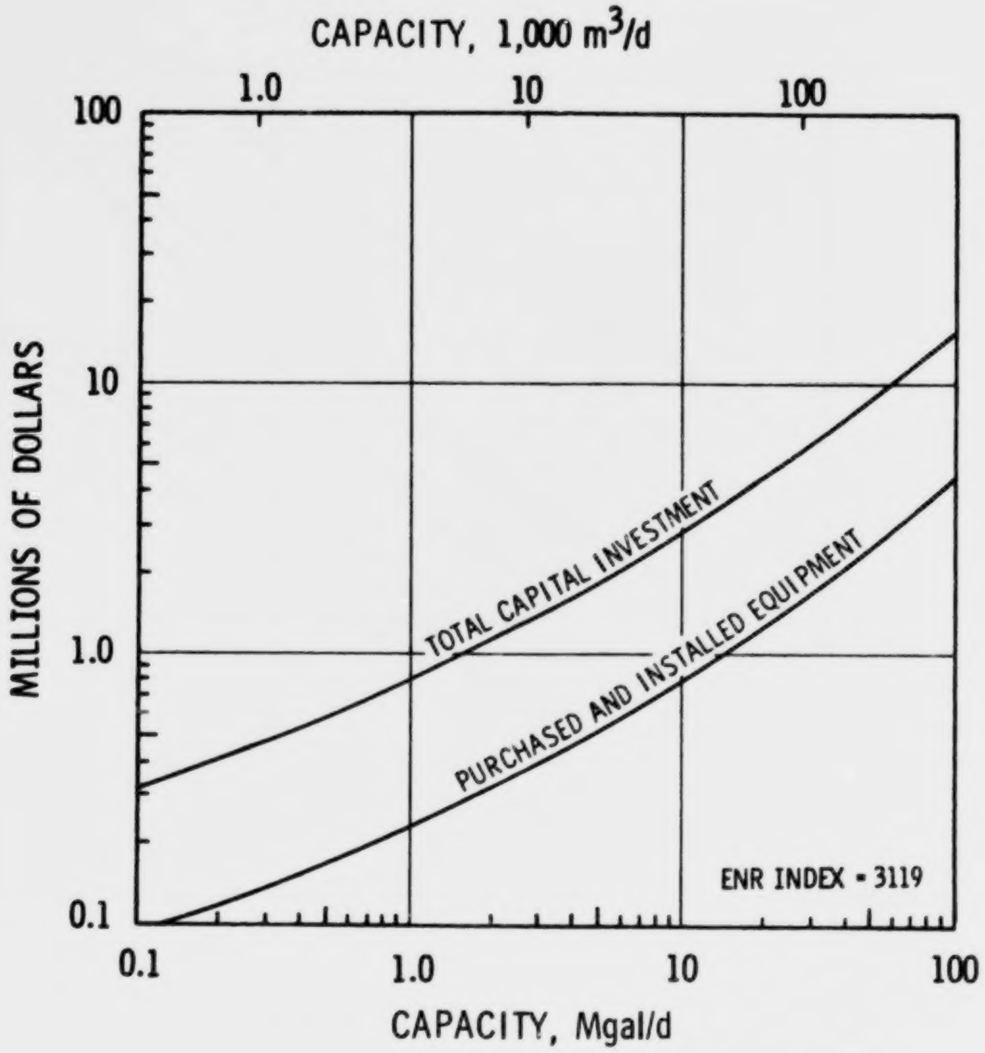
IV.2.3.3 Costs

Purchased equipment and installation cost is based on a concrete basin for design flows less than 1 Mgal/d and 6-in. concrete-lined earthen basin for design flows greater than 1 Mgal/d. Detention time equals 1.0 day. Capital costs include mixers. Operating costs are based on mechanical mixing at 20 hp/Mg and pumping at 10 ft TDH.

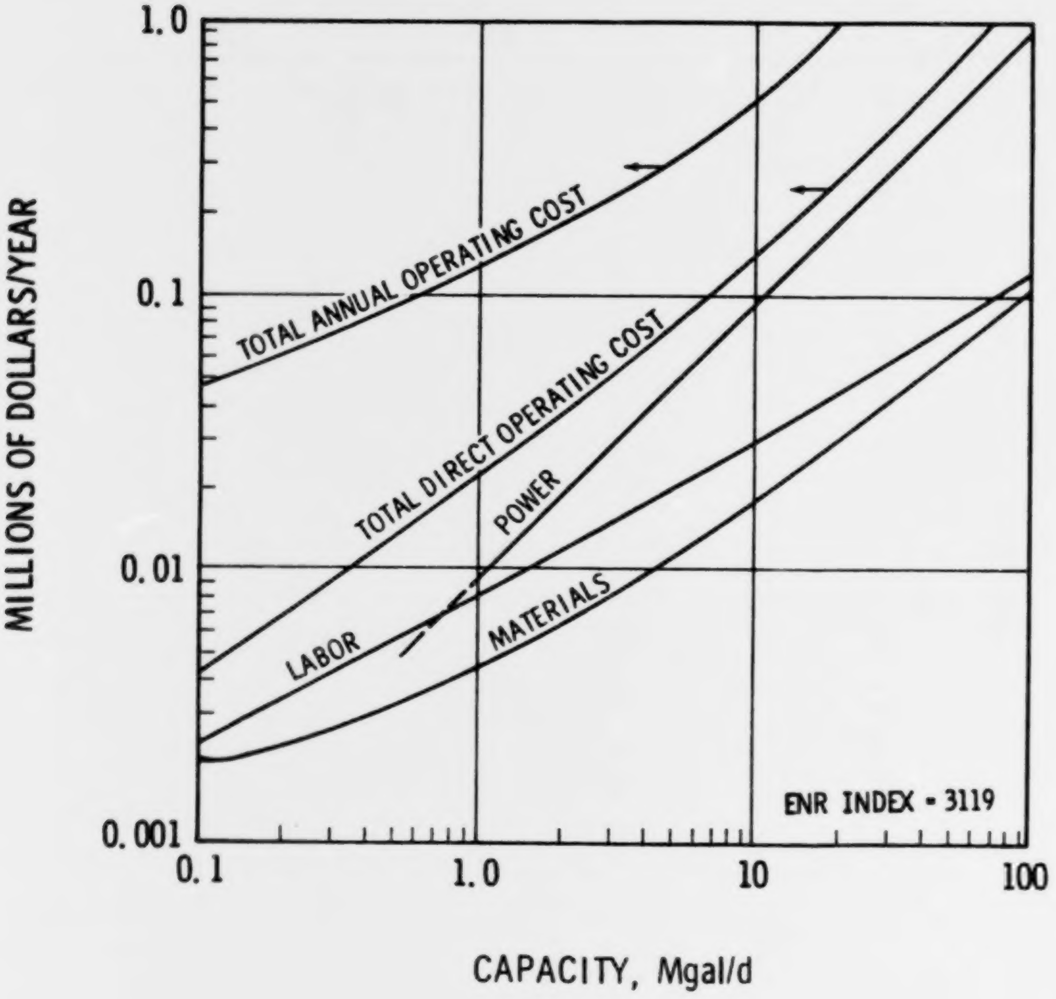
IV.2.3.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.2.4 NEUTRALIZATION [1]

IV.2.4.1 Description

Neutralization is the process of adjusting either an acidic or a basic wastestream to a pH in the range of seven. Since many manufacturing and processing operations produce effluents that are acidic or alkaline in nature, neutralization of an excessively acidic or basic waste stream is necessary in a variety of situations. For example it is required: (1) to prevent metal corrosion and/or damage to other construction materials; (2) to protect aquatic life and human welfare; (3) as a preliminary treatment allowing effective operation of biological treatment processes, and (4) to provide neutral pH water for recycle, either as process water or as boiler feed. Treatment to adjust pH may also be desirable to break emulsions, to insolubilize certain chemical species, or to control chemical reaction rates, e.g., chlorination. Although natural waters may differ widely in pH, changes in a particular pH level could produce deterimental effects on the environment. To minimize any undesirable consequences, the effluent limitations guidelines for industrial sources set the pH limits for most industries between 6.0 and 9.0 for 1977 and 1983.

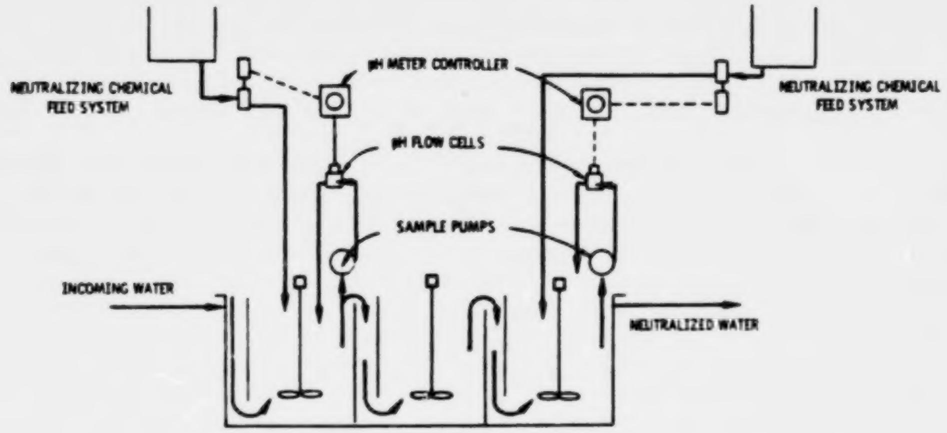
Simply, the process of neutralization is the interaction of an acid with a base. The typical properties exhibited by acids in solution are due to the hydrogen ion, (H^+). Similarly, alkaline (or basic) properties are a result of the hydroxyl ion, (OH^-). In aqueous solutions, acidity and alkalinity are defined with respect to pH, where $pH = - \log (H^+)$ and, at room temperature, $pH = 14 + \log (OH^-)$. In the strict sense, neutralization is the adjustment of pH to 7, the level at which the concentrations of hydrogen and hydroxyl ions are equal. Solution with excess hydroxyl ion concentration (pH >7) are said to be basic, solutions with excess hydrogen ions (pH <7) are acidic. Since adjustment of the pH to 7 is not often practical or even desirable in waste treatment, the term "neutralization" is sometimes used to describe adjustment of pH to values other than 7.

The actual process of neutralization is accomplished by the addition of an alkaline to an acidic material or by adding an acidic to an alkaline material, as determined by the required final pH. The primary products of the reaction are a salt and water. A simple example of acid-base neutralization is the reaction between hydrochloric acid and sodium hydroxide:

$$HCl + NaOH \rightarrow H_2O + NaCl$$

The product, sodium chloride in aqueous solution, is a neutral with pH = 7.0.

Required equipment may be as varied as the reactions that can be classified as neutralizations. However, the most commonly occurring scheme for neutralization is the addition of a chemical, in either liquid or solid form, directly into a wastewater stream. The required equipment for this treatment is simple, i.e., storage and reaction tanks with accessory agitators, and delivery systems. The tanks may be of any shape, round or square, but must be properly baffled to allow adequate mixing and prevent "short-circuiting." Frequently the neutralization is carried out in a series of reactors to provide better control of the final pH. Such a system with three units is shown in the flow diagram below. There are two addition points for neutralizing agent, controlled by pH monitors. Most of the neutralization occurs in the first reactor, adjustment to the final desired pH takes place just before discharge. In large-scale operations the tanks are often built below ground level and constructed of concrete.


NEUTRALIZING CHEMICAL FEED SYSTEM
pH METER CONTROLLER
NEUTRALIZING CHEMICAL FEED SYSTEM
pH FLOW CELLS
SAMPLE PUMPS
INCOMING WATER
NEUTRALIZED WATER

IV.2.4.2 Typical Design Criteria

The design of a neutralization system is highly specific. The following design criteria is based on the requirements for lime neutralization of sulfuric acid waste stream and is discussed in the following section.

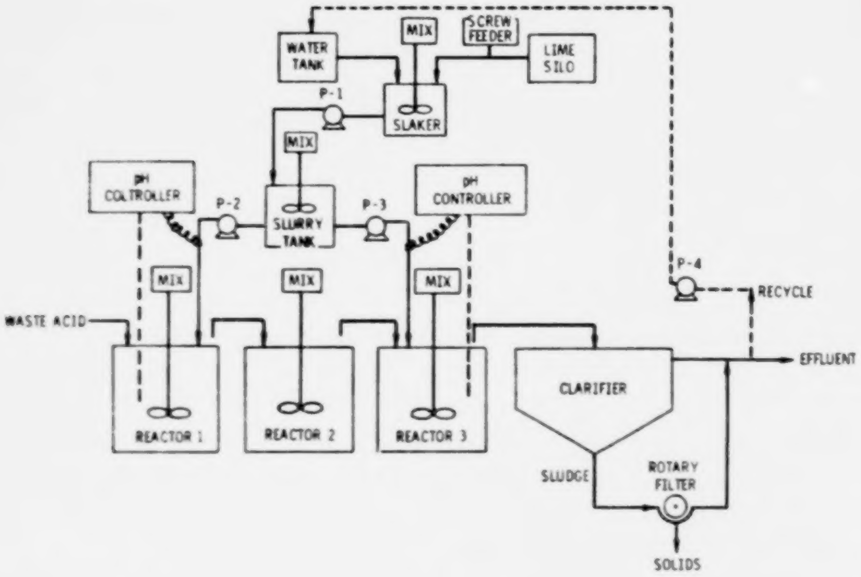
Design criteria	Value
Basis	1 Mgal/d; 1% H_2SO_4 (by vol); 350 day/yr operation
Equipment	
Three reactors[a] (with mixers)	7,500 gal each
Slaking tank (with mixer)	1,000 gal
Lime slurry tank (with mixer)	5,000 gal
Water storage tank	200,000 gal
Lime/silo	2,000 ft^3
Clarifier	250,000 gal
Rotary filter	11,250 lb/hr dry solids (filter cake 50-60% solids)
Lime screw feeder	3,700 lb/hr
Pumps	
P-1	70 gpm
P-2	70 gpm
P-3	35 gpm
P-4	70 gpm
Two pH recorder/controllers (with sample pumps)	
Chemicals	
Quicklime	44 ton/d
Power	
Electricity (annual basis)	675,000 kWh
Labor	32 man-hr/d

[a] Due to possible acidic conditions, reactors should be lead- or rubber-lined; agitator shafts and blades should be rubber-coated.

IV.2.4.3 Costs

The capital investment for neutralization depends on the specific treatment process selected and the size of the stream to be treated, and it can result in differences in capital costs of orders of magnitude. For example, a 1971 study of limestone treatment of acid mine drainage (sulfuric acid) estimated that capital costs for treating a stream of about 900 gpm could range from $352,000 to $1,238,000, depending on the composition of the stream. High levels of iron, aluminum, and sulfate contribute to larger capital investment, particularly in the area of sludge handling equipment. For the same reason the estimated cost of operation ranged from $0.35 to $3.76 per 1,000 gal (including sludge disposal). This example indicates the potentially large variation in investment and operating costs in the same process operating on different quality wastes. It would be inappropriate to draw conclusions concerning investment capital for neutralization without some knowledge of the system. Materials of construction and the need for sludge-handling equipment are just two additional factors that can drastically alter an economic analysis. With regard to operating costs, similar considerations, such as sludge disposal costs, may be critical.

To provide a general understanding of the economics of a neutralization process a "simulated" sample system is illustrated in the figure below. This design, along with the description in the following table, represents a common scheme for lime neutralization of a waste stream containing sulfuric acid. An economic analysis for the treatment process also is shown below.


WATER TANK
MIX
SCREW FEEDER
LIME SILO
P-1
SLAKER
MIX
pH CONTROLLER
P-2
SLURRY TANK
P-3
pH CONTROLLER
P-4
RECYCLE
MIX
MIX
MIX
WASTE ACID
EFFLUENT
CLARIFIER
REACTOR 1
REACTOR 2
REACTOR 3
SLUDGE
ROTARY FILTER
SOLIDS

Placing the exact price on the sludge disposal is difficult. The chemical composition of the material will influence the choice of disposal method. A sludge containing large quantities of heavy metals may not be suitable for landfill, due to the possibility of leaching, and a more costly method may have to be devised. Conversely, a relatively pure sludge of calcium sulfate might be sold as a by product and count as a credit.

Fixed capital cost based on an ENR index of 3119 is estimated to be $1,420,000. Similarly, total capital cost is estimated to be $1,504,000. Estimation of annual operating cost is presented in the following table.

ESTIMATE OF ANNUAL OPERATING COST FOR LIME NEUTRALIZATION OF SULFURIC ACID WASTE STREAM

Cost item	Annual quantity	Cost per unit quantity	Annual cost,[a] $	
Direct operating cost				
Labor				
Operating	11,200 man-hr	$16/hr	180,000	
Maintenance			28,000	
				208,000
Chemicals				
Lime	15,400 tons	$35/ton	539,000	
Coagulant aids	3,000 lb	$2.00/lb	6,000	
				545,000
Materials - Maintenance				29,000
Power	675,000 kWh	$0.035/kWh		23,600
Total				806,000
Total indirect operating cost				389,000
Total annual operating cost				1,200,000

[a] Costs have not been included for sludge disposal or, conversely for gypsum credit, if the gypsum is saleable.

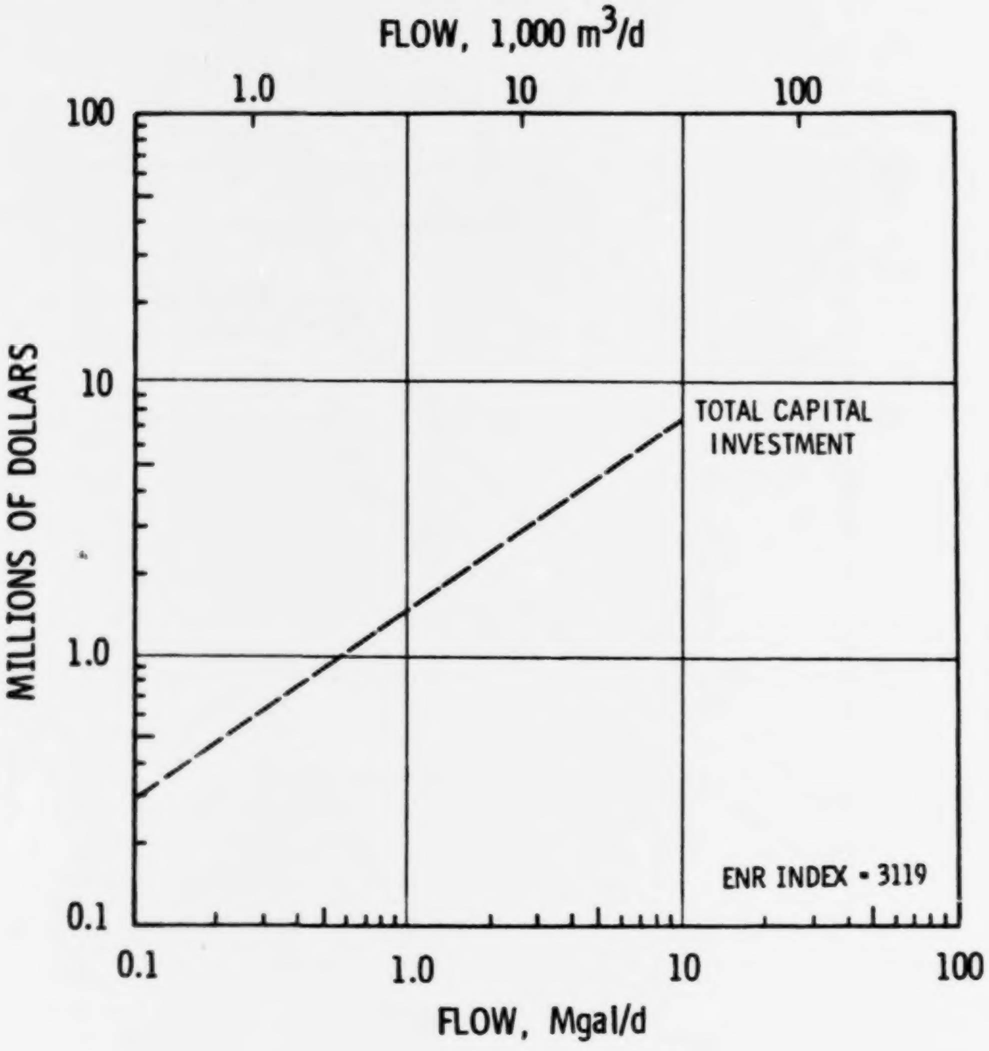
Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying capacities:

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Materials, power and chemicals	1.0

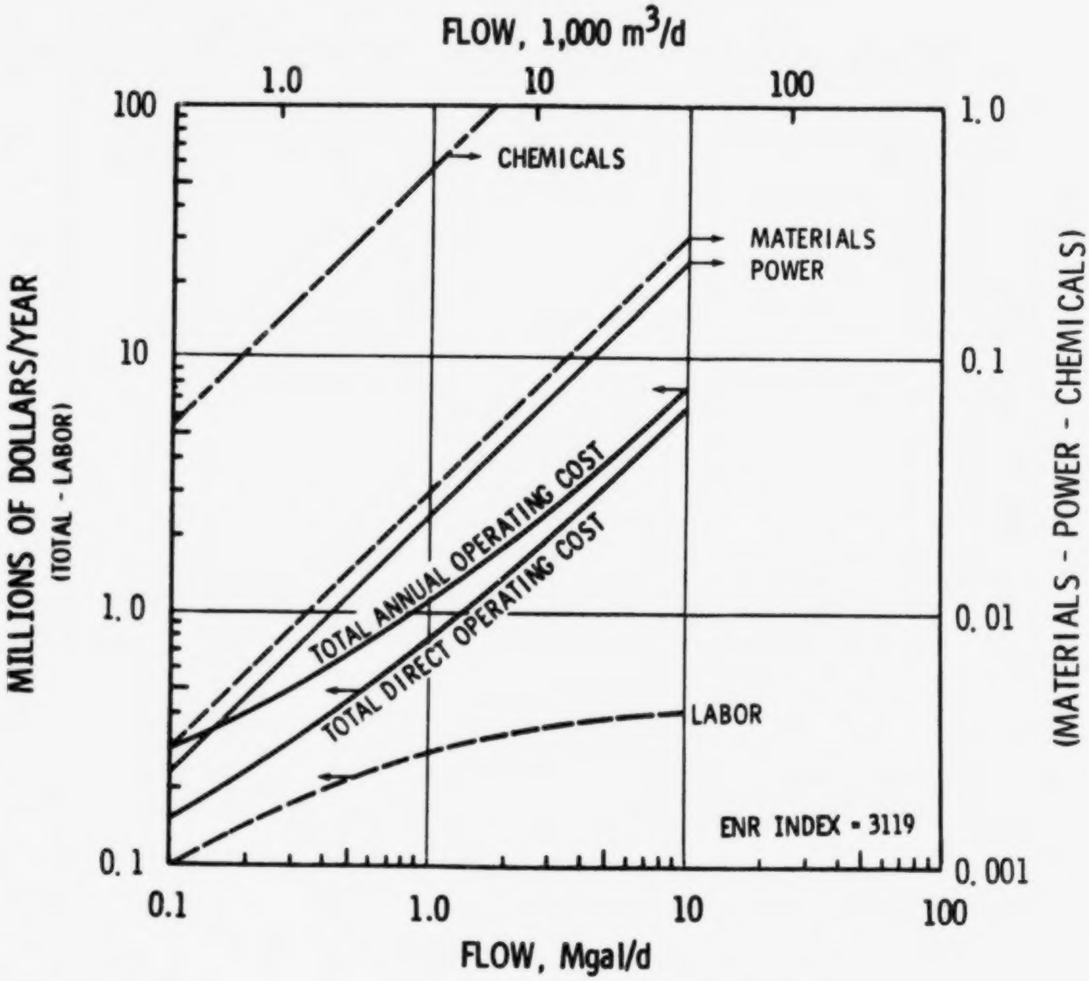
IV.2.4.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 34-1 to 34-20.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3 PRIMARY WASTEWATER TREATMENT

IV.3.1 GRAVITY OIL SEPARATION

IV.3.1.1 Description

Gravity oil separation is used for the removal of floatable oil and grease.

A gravity oil separator (skimming tank) is a chamber so arranged that floating matter rises and remains on the surface of the wastewater until removed, while the liquid flows out continuously through deep outlets or under partitions, curtain walls, or deep scum boards. This may be accomplished in a separate tank or combined with primary sedimentation, depending on the process and nature of the wastewater.

The objective of skimming tanks is the separation from the wastewater of the lighter floating substances. The material collected on the surface of skimming tanks, whence it can be removed, may include oil, grease, soap, pieces of cork and wood, vegetable debris, and fruit skins. The outlet, which is submerged, is opposite the inlet and at a lower elevation to assist in flotation and to remove any solids that may settle.

Gravity-type separators are the most common devices employed in oily waste treatment. The effectiveness of a gravity separator depends upon proper hydraulic design, and the design period of wastewater retention. Longer retention times allow better separation of the floatable oils from the water. Short detention times of less than 20 minutes result in less than 50% oil-water separation, while more extended holding periods improve oil separation from the waste stream.

Gravity separators are equally effective in removing both greases and nonemulsified oils. The standard unit in refinery waste treatment is the API separator, based upon design standards published by the American Petroleum Institute. Separators used for metal and food processing oily wastes operate upon the same principle of floating the oil, and many are designed in a similar fashion to the API process insofar as skimming, retention time, etc. Separators may be operated as batch vats, or as continuous flow-through basins, depending upon the volume of waste to be treated.

IV.3.1.2 Typical Design Criteria

Design criteria	Value
Surface loading[a]	$\leq$700 gpm/ft²
Wastewater depth	6-7 ft
Detention time	1-2 hr

[a] Many states require the surface loading to be $\leq$700 gpm/ft².

To adjust costs for alternative surface overflow rate, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{800 \text{ gpd/ft}^2}{\text{new design surface overflow rate}}$$

IV.3.1.3 Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps and separator. The following operating characteristics were assumed for cost estimation.

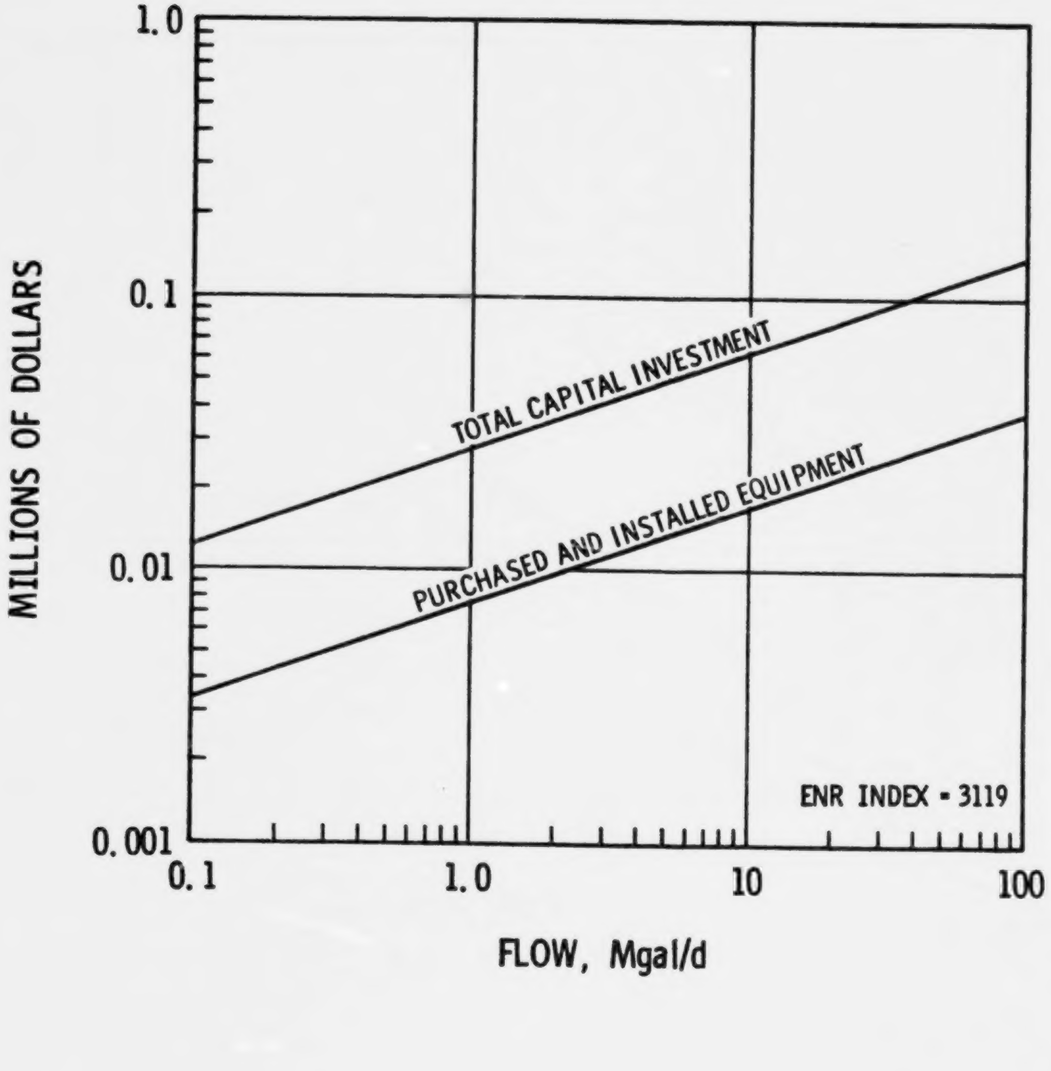
Operating characteristic	Assumed value
Service life	50 yr
Surface overlow rate	800 gpd/ft² (based on average flow)
Sludge concentration	4% solids
Pump head	10 ft TDH

IV.3.1.4 References

1. Patterson, J. W. Wastewater Treatment Technology. Ann Arbor Science Publishers, Inc., Ann Arbor, Michigan, 1975. pp. 179-185.

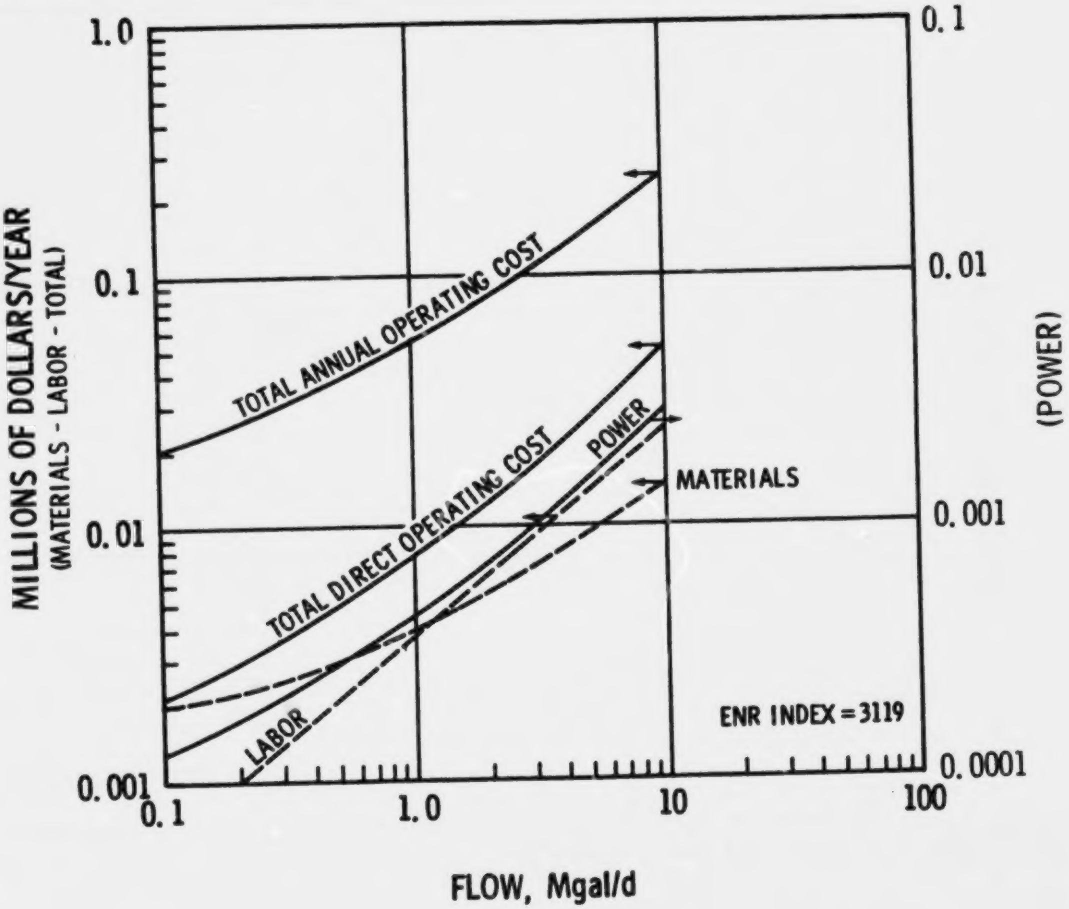
2. Azad, H. S. Industrial Wastewater Management Handbook, McGraw Hill Book Company, New York, 1976. p. 5-28.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



IV.3.2 CLARIFICATION/SEDIMENTATION

IV.3.2.1 Primary Circular Clarification [1]

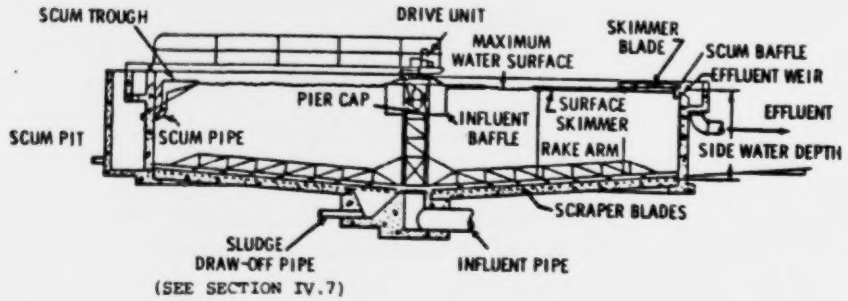
Description. Primary circular clarification involves a relatively long period of quiescence in a basin (depths of 10 to 15 feet) where most of the settleable solids fall out of suspension by gravity; a chemical coagulant may be added. The solids are mechanically collected on the bottom and pumped as a sludge underflow.

The conical bottom (one inch per foot of slope) is equipped with a rotating mechanical scraper that plows sludge to a center hopper. An influent feed well located in the center distributes the influent radially, and a peripheral weir overflow system carries the effluent. Floating scum is trapped inside a peripheral scum baffle and squeegeed into a scum discharge box. The unit contains a center motor-driven turntable drive supported by a bridge spanning the top of the tank, or supported by a vertical steel center pier. The turntable gear rotates a vertical cage or torque tube, which in turn rotates the truss arms (preferably two long arms). The truss arms carry multiple flights (plows) on the bottom chord that are set at a 30° angle of attack and literally "plow" heavy fractions of sludge and grit along the bottom slope toward the center blowdown hopper. An inner diffusion chamber receives influent flow and distributes it (by means of about a four-inch water head loss) inside of the large diameter feedwell skirt.

Approximately three percent of the clarifer surface area is used for the feed-well. The depth of the feed-wells is generally about one-half of the tank depth. The center sludge hopper should be less than two feet deep and less than four square feet in cross section.

Two short auxiliary scraper arms are added perpendicular to the two long arms on medium-to-large tanks. This makes practicable the use of deep spiral flights, which aid in center region plowing where ordinary shallow straight plows (30° angle of attack) are nearly useless. Peripheral feed systems are sometimes used in lieu of central feed. Also, central effluent weirs are sometimes used. Flocculating feed wells also may be provided if coagulants are to be added to assist sedimentation.

Equipment normally associated with primary circular clarification includes a clarifier and sludge pumps. A flow diagram for primary circular clarification is shown on the following page.


SCUM TROUGH
DRIVE UNIT
MAXIMUM WATER SURFACE
SKIMMER BLADE
SCUM BAFFLE
EFFLUENT WEIR
EFFLUENT
PIER CAP
INFLUENT BAFFLE
SURFACE SKIMMER
RAKE ARM
SCUM PIT
SCUM PIPE
SIDE WATER DEPTH
SCRAPER BLADES
SLUDGE DRAW-OFF PIPE
(SEE SECTION IV.7)
INFLUENT PIPE

Typical design criteria. Typical design criteria for a primary circular clarification system are tabulated below.

Design criteria	Value
Surface loading	600 - 1,200 gpd/ft²
Detention time	1.5 - 3.0 hr
Weir loading	10,000 - 70,000 gpd/linear ft
Sludge pumping rate[a]	2,500 - 20,000 gpd/Mgal
Domestic scum handling requirement	6 ft³/Mgal decant water
Sludge collector tip speed	10 - 15 ft/min

[a] Depends on chemical addition and service.

Additionally, heads of 2 to 3 ft H_2O are required to overcome losses at inlet and effluent controls and in connecting pipes. Forward velocity should be less than 9 to 15 times the particle settling velocity to avoid scour.

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps and clarifiers. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Service life	50 yr
Surface overflow rate	800 gpd/ft² (based on average flow)
Sludge concentration	4% solids
Pump head	10 ft TDH

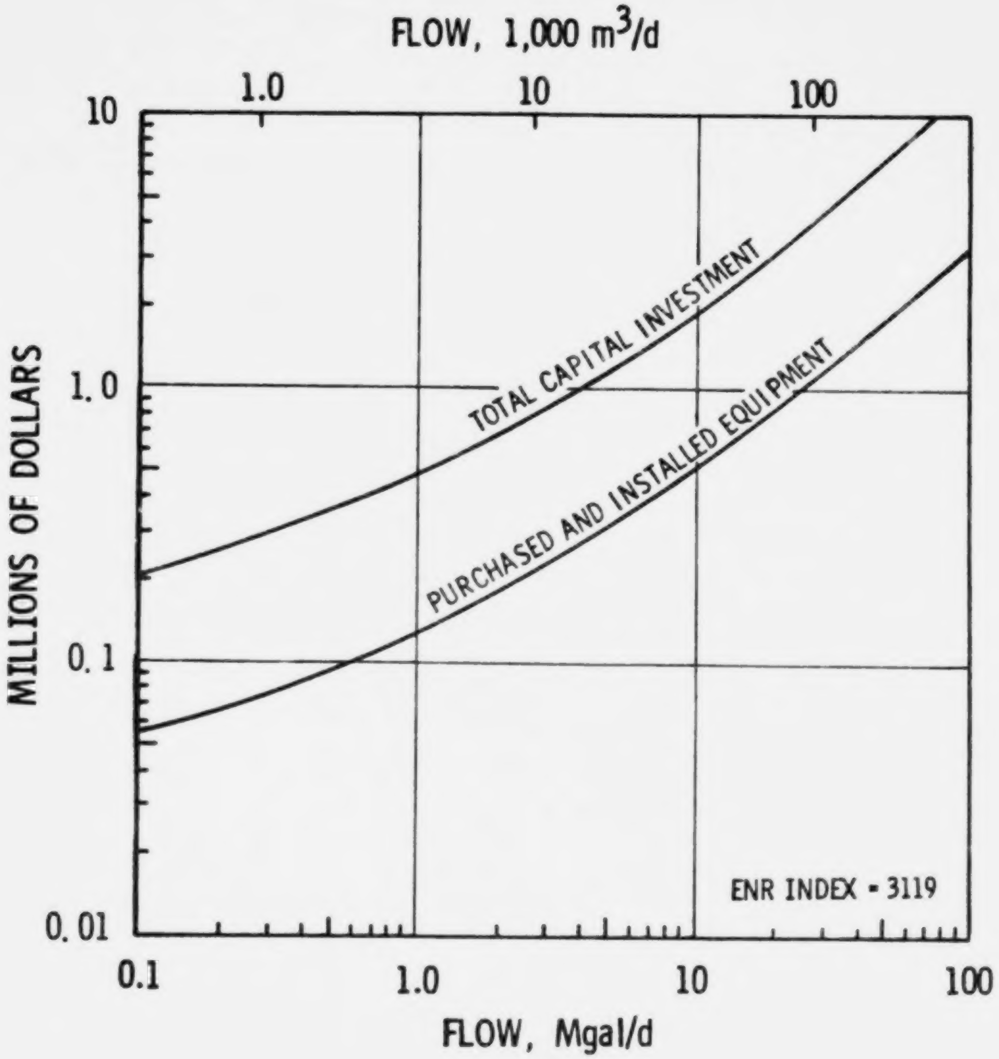
To adjust costs for alternative surface overflow rate, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{800\ \text{gpd/ft}^2}{\text{new design surface overflow rate}}$$

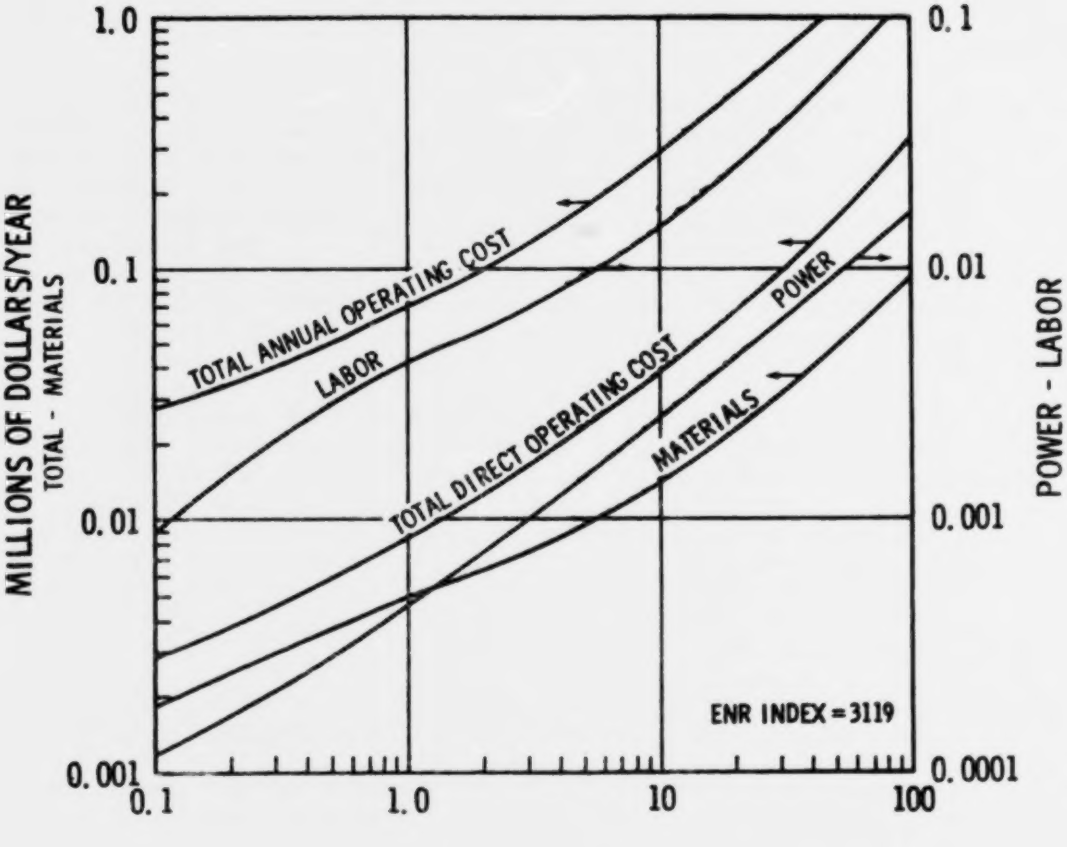
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.2.2 Primary Rectangular Clarification [1]

Description. Primary rectangular clarification involves a relatively long period of quiescence in a basin (depths of 10 to 15 feet) where most of the settleable solids in the wastewater fall out of suspension by gravity. The solids are mechanically transported along the bottom of the tank by a scraper mechanism and pumped as a sludge underflow.

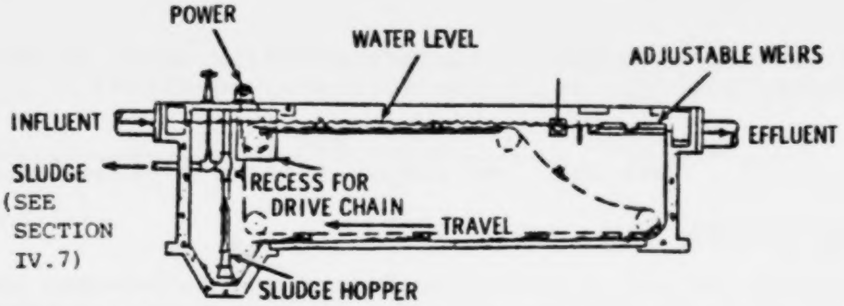
The maximum length of rectangular tanks has been approximately 300 feet. Where widths greater than 20 feet are required, multiple bays with individual cleaning equipment may be employed, thus permitting tank widths up to 80 feet or more. Influent channels and effluent channels should be located at opposite ends of the tank.

Sludge removal equipment usually consists of a pair of endless conveyor chains. Attached to the chains at about 10 foot intervals are wooden crosspieces or flights, extending the full width of the tank or bay. Linear conveyor speeds of 2 to 4 ft/min are common. The settled solids are scraped to sludge hoppers in small tanks and to transverse troughs in large tanks. The troughs, in turn, are equipped with cross collectors, usually of the same type as the longitudinal collectors, which convey solids to one or more sludge hoppers. Screw conveyors have been used for the cross collectors.

Scum is usually collected at the effluent end of rectangular tanks by the flights returning at the liquid surface. The scum is moved by the flights to a point where it is trapped by baffles before removal, or it can be moved along the surface by water sprays. The scum is then scraped manually up an inclined apron, or it can be removed hydraulically or mechanically, and for this process a number of means have been developed (rotating slotted pipe, transverse rotating helical wiper, chain and flight collectors, scum rakes).

Tanks may be cleaned by a bridge-type mechanism which travels up and down the tank on rails supported on the sidewalls. Scraper blades are suspended from the bridge and are lifted clear of the sludge on the return travel. Chemical coagulants may be added to improve BOD_5 and suspended solid removals and to remove phosphorus ion.

Equipment normally associated with primary rectangular clarification includes clarifiers and sludge pumps. A flow diagram for primary rectangular clarification is shown on the following page.


POWER
WATER LEVEL
ADJUSTABLE WEIRS
INFLUENT
EFFLUENT
SLUDGE
(SEE SECTION IV.7)
RECESS FOR DRIVE CHAIN
TRAVEL
SLUDGE HOPPER

Typical design criteria. Typical design criteria for a primary rectangular clarification system are tabulated below.

Design criteria	Values
Surface loading[a]	600 - 1,200 gpd/ft²
Detention time	1.5 - 3.0 hr
Weir loading	10,000 - 30,000 gpd/linear ft
Length/width ratio	At least 4/1 per day
Sludge pumping rate[b]	2,500 - 20,000 gpd/Mgal
Scum handling requirement	6 ft³/Mgal decant water

[a] Average for untreated wastewater. If chemicals are used, the ranges are 360 to 600 gpd/ft² for alum, 540 to 800 gpd/ft² for iron, and 1600 gpd/ft² for lime.

[b] Depends on chemical addition and service.

Additionally, forward velocities should be less than 9 to 15 times the settling velocity to avoid scour.

Costs. Purchased equipment and installation costs for estimation of the total capital investment includes primary sludge pumps and clarifiers. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Service life	50 yr
Sludge concentration	4% solids
Surface overflow rate	800 gpd/ft²
Pump head	10 ft TDH

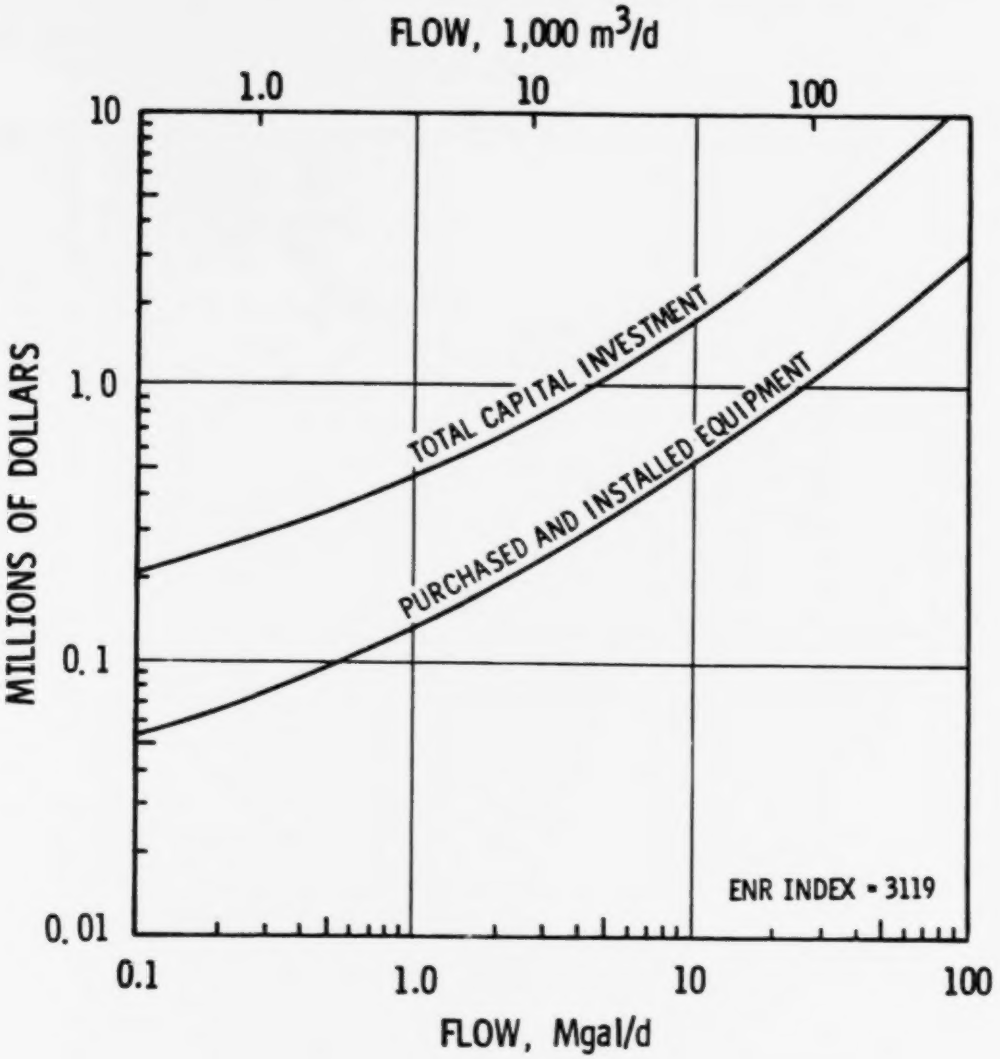
To adjust for alternative surface overflow rate, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{800 \text{ gpd/ft}^2}{\text{new design surface overflow rate}}$$

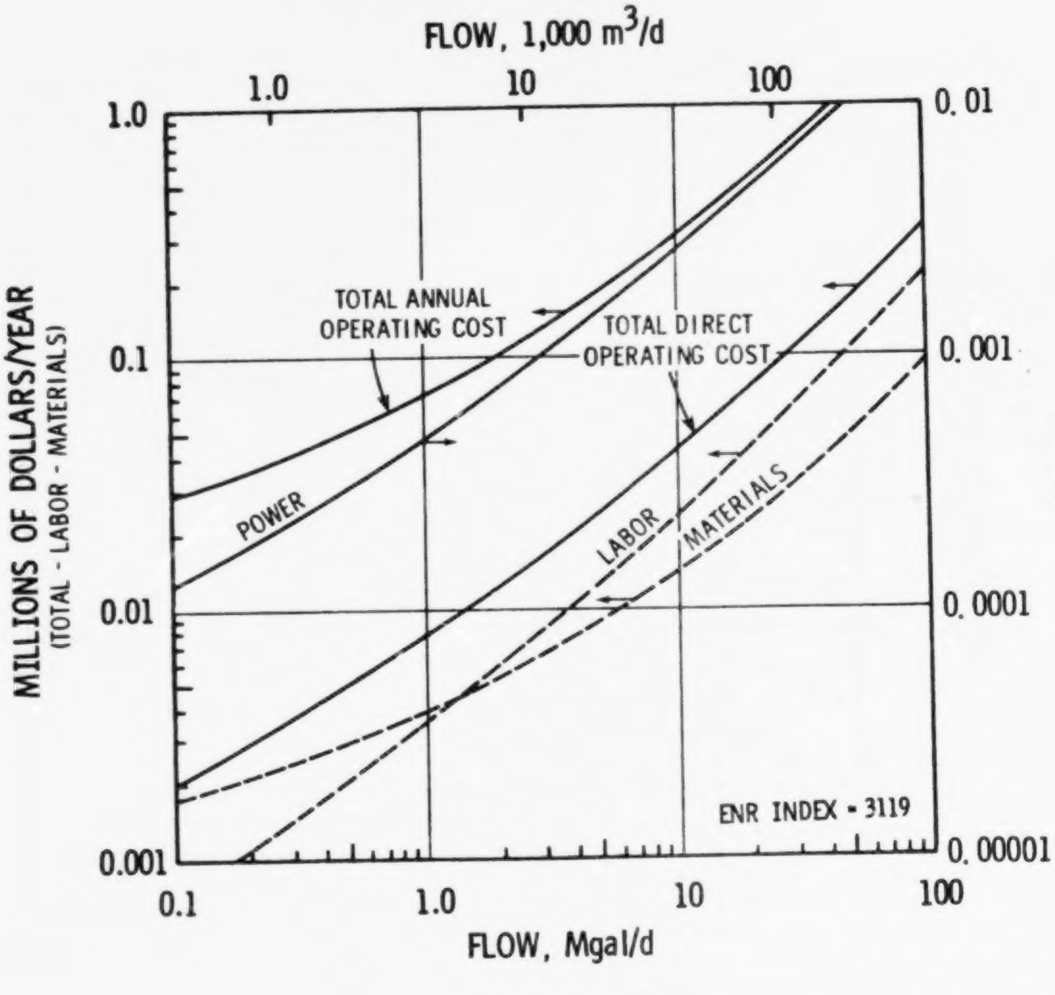
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.2.3 Secondary Circular Clarification [1]

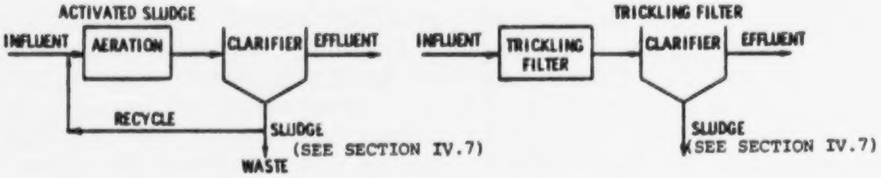
Description. Secondary circular clarifiers have been constructed with diameters ranging from 12 to 200 feet with depths of 12 to 15 feet. There are two basic types: the center-feed and the rim-feed. Both utilize a revolving mechanism to transport and remove the sludge from the bottom of the clarifier. Mechanisms are of two types: those that scrape or plow the sludge to a center hopper similar to the types used in primary sedimentation tanks, and those that remove the sludge directly from the tank bottom through suction orifices that serve the entire bottom of the tank in each revolution. In one of the latter, the suction is maintained by reduced static head on the individual drawoff pipes. In another patented suction system, sludge is removed through a manifold either hydrostatically or by pumping.

Secondary circular clarifiers are made with effluent overflow weirs located near the center or near the perimeter of the tank. Skimming facilities are required on all federally funded projects.

While the design is similar to primary clarifiers, the large volume of flocculent solids in the mixed liquor requires that special consideration be given to the design of activated sludge clarifiers. The sludge pump capacity and the size of the settling tank are larger. Further, the mixed liquor, on entering the tank, has a tendency to flow as a density current interfering with the separation of the solids and the thickening of the sludge. To cope successfully with these characteristics, the following factors must be considered in the design of these tanks: type of tank to be used, surface loading rate, solids loading rate, flow-through velocities, weir placement and loading rates, and scum removal.

Multiple inlets are used with balanced flow at various spacings with target baffles to reduce velocity of streams. Hydraulic balancing is used between parallel clarifier units. Wind effects on water surface are controlled. Sludge hopper collection systems and flocculation inlet structures are used. Traveling bridge sludge collectors and skimmers are used an an alternate to chain and flight systems.

Equipment normally associated with secondary circular clarification includes clarifiers and sludge pumps. Flow diagrams of secondary circular clarification systems are shown on the next page.


ACTIVATED SLUDGE
INFLUENT
AERATION
CLARIFIER
EFFLUENT
RECYCLE
SLUDGE
(SEE SECTION IV.7)
WASTE
TRICKLING FILTER
INFLUENT
TRICKLING FILTER
CLARIFIER
EFFLUENT
SLUDGE
(SEE SECTION IV.7)

Typical design criteria. Typical design criteria for a secondary circular clarification system are tabulated below.

Design criteria	Secondary circular clarification air activated sludge[a]	Trickling filtration	Air activated sludge	Extended aeration	Secondary clarification following oxygen activated sludge	Units
Hydraulic loading (average)	400 - 800	400 - 600	400 - 800	200 - 400	400 - 800	gpd/ft²
Hydraulic loading (peak)	700 - 1,200	1,000 - 1,200	1,000 - 2,000	800	1,000 - 1,200	gpd/ft²
Solids loading (average)	14 - 28	-	20 - 30	20 - 30	25 - 35	lb/d/ft²
Solids loading (peak)	30 - 48	-	50	50	50	lb/d/ft²
Depth	12 - 15	10 - 12	12 - 15	12 - 15	12 - 15	ft
Inlet baffle diameter	-	15 - 30	15 - 20	15 - 20	15 - 20	Percent of tank diameter
Weir loading rate	10,000 - 30,000	10,000 - 30,000	10,000 - 30,000	10,000 - 30,000	10,000 - 30,000	gpd/linear ft
Maximum upflow velocity in vicinity of weir	12 - 36	12 - 36	12 - 36	12 - 36	12 - 24	ft/hr

[a] Excluding extended aeration.

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes sludge return and waste pumps. Spare pumps (nonclog centrifugal) are included as necessary. The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Service life	40 yr
Clarifier type	Flocculator-type
Overflow rate	600 gpd/ft²
Sludge concentration	1% solids
Pump head	10 ft TDH
Pump capacity	350 gpm/Mgpd of plant capacity
Maximum clarifier diameter	200 ft

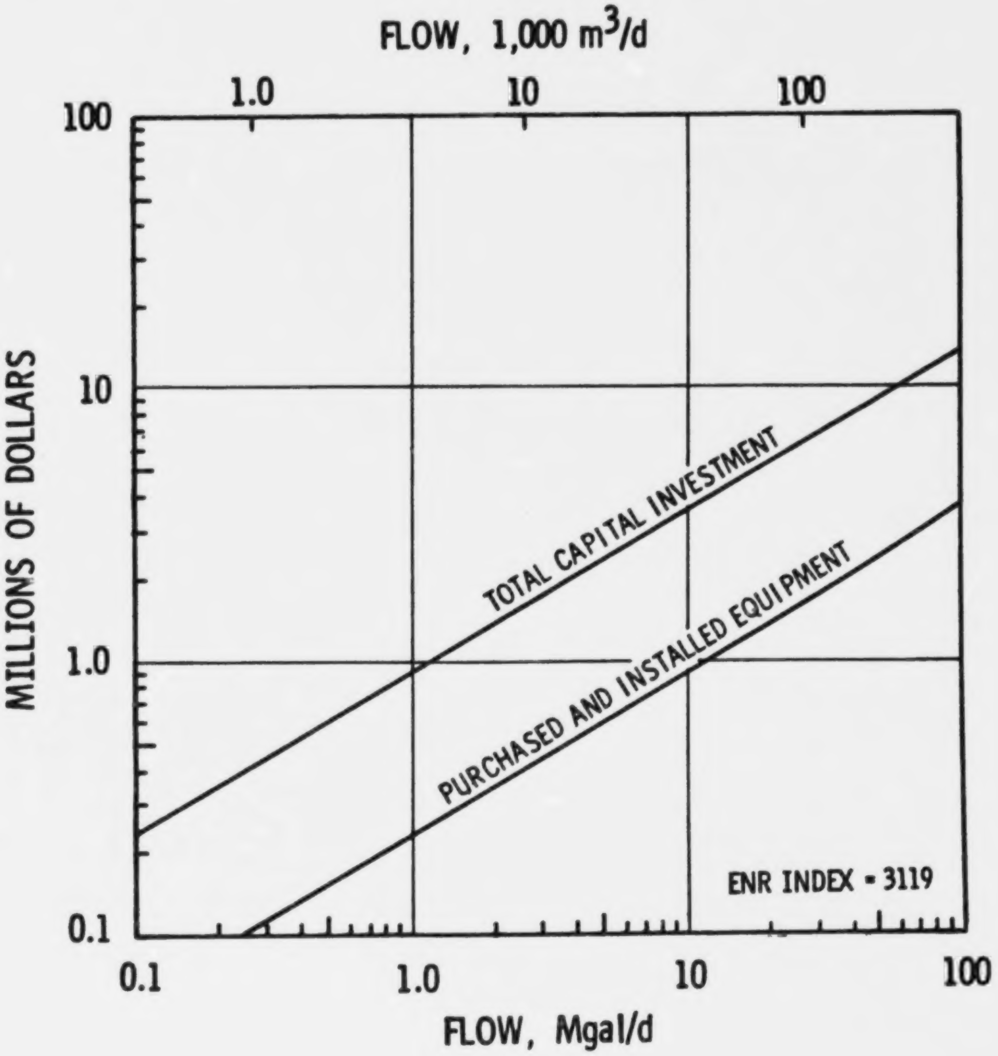
To adjust costs for an alternative overflow rate, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{600 \text{ gpd/ft}^2}{\text{new design overflow rate}}$$

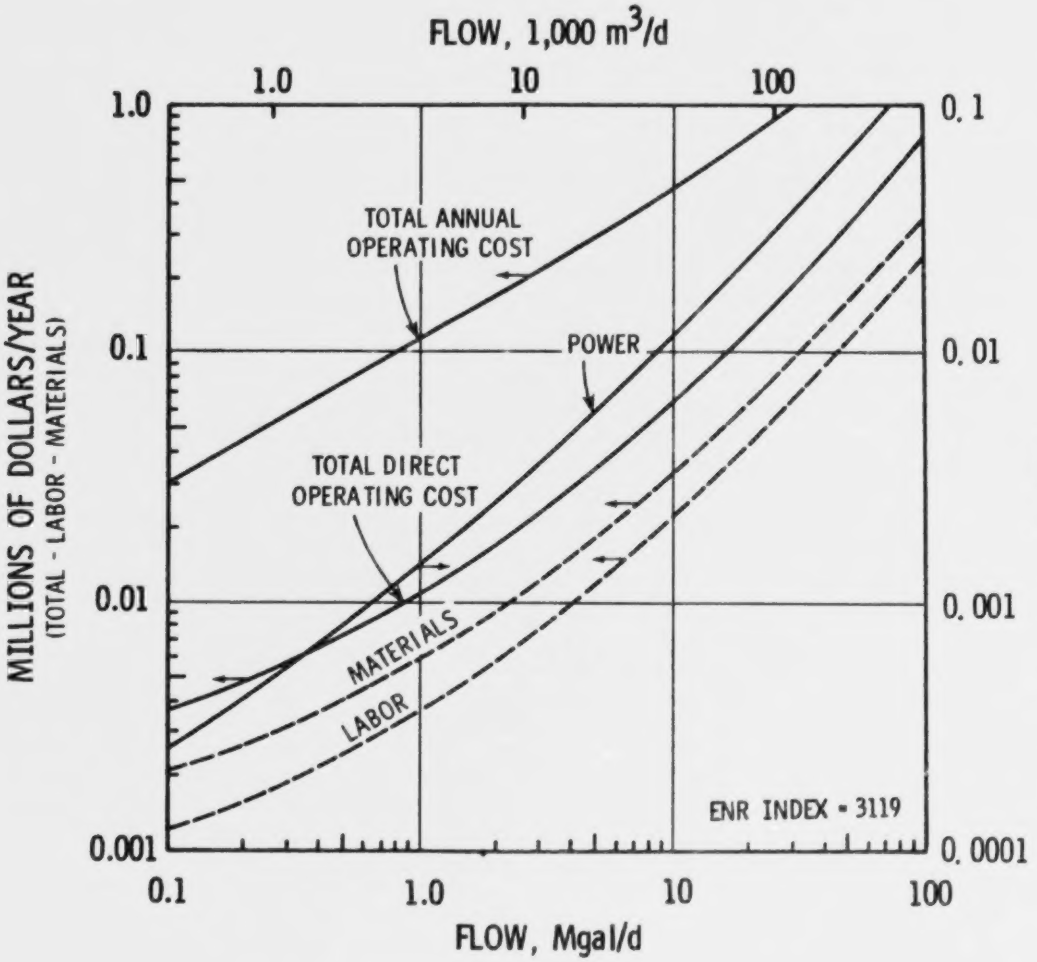
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.2.4 Secondary Rectangular Clarification [1]

Description. The design of secondary rectangular clarifiers is similar to that of primary rectangular clarifiers except that the large volume of flocculent solids in the mixed liquor must be considered during the design of activated sludge clarifiers and in sizing the sludge pumps. Further, the mixed liquor, on entering the tank, has a tendency to flow as a density current interfering with the separation of the solids and the thickening of the sludge. To cope successfully with these characteristics, the following factors must be considered in the design of these tanks: (1) type of tank to be used, (2) surface loading rate, (3) solids loading rate, (4) flowthrough velocities, (5) weir placement and loading rates, and (6) scum removal.

In rectangular tanks, the flow enters at one end, passes a baffle arrangement, and traverses the length of the tank to the effluent weirs. The maximum length of rectangular tanks has been approximately 300 feet with depths fo 12 to 15 feet. Where widths greater than 20 feet are required, multiple bays with individual cleaning equipment may be employed, thus permitting tank widths up to 80 feet or more.

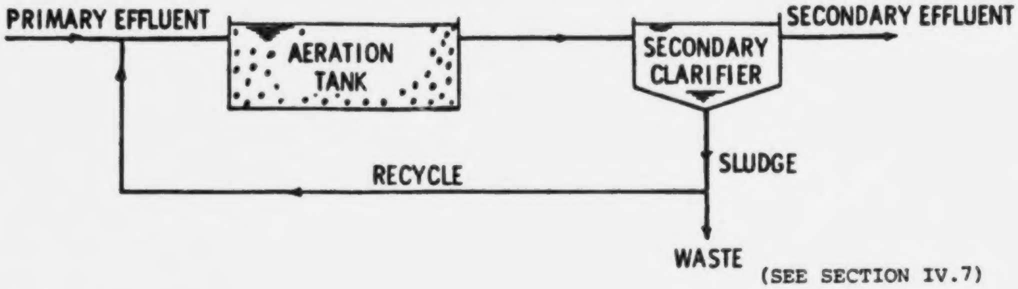
Sludge removal equipment usually consists of a pair of endless conveyor chains. Attached to the chains at 10 foot intervals are 2-inch thick wooden crosspieces or flights, 6 to 8 inches deep, extending the full width of the tank or bay. Linear conveyor speeds of 2 to 4 ft/min are common. The settled solids are scraped to sludge hoppers in small tanks and to transverse troughs in large tanks. The troughs, in turn, are equipped with cross collectors, usually of the same type as the longitudinal collectors, which convey solids to one or more sludge hoppers. Conveyors also have been used for the cross collection.

Tanks also may be cleaned by a bridge-type mechanism that travels up and down the tank on rails supported on the sidewalls. Scraper blades are suspended from the bridge and are lifted clear of the sludge on the return travel. For very long tanks, it is desirable to use two sets of chains and flights in tandem with a central hopper to receive the sludge. Tanks in which mechanisms that move the sludge toward the effluent end in the same direction as the density current have shown superior performance in some instances.

Scum is usually collected at the effluent end of rectangular tanks by the flights returning at the liquid surface. The scum is moved by the flights to a point where it is trapped by baffles before removal, or it can be moved along the surface by water sprays. The scum is then scraped manually up an inclined apron, or it can be removed hydraulically or mechanically, and for this process a number of means have been developed (rotating slotted pipe, transverse rotating helical wiper, chain and flight collectors, scum rakes).

Common modifications of secondary rectangular clarification include multiple inlets with balanced flow at various spacings with target baffles to reduce velocity of streams; hydraulic balancing between parallel clarifier units; control of wind effects on water surfaces; sludge hopper collection systems; flocculation inlet structures; use of traveling bridge sludge collectors and skimmers as an alternate to chain and flight systems; use of steeply inclined tube settlers to enhance suspended solids removal in either new or rehabilitated clarifiers; use of wedge wire settler panels at peak hydraulic loading of less than 800 gpd/ft^2 for improved suspended solids removal.

A flow diagram of secondary rectangular clarification is shown below.



Typical design criteria. Typical design criteria for secondary rectangular clarification are shown below.

Design criteria	Value
Average hydraulic loading[a]	400 - 800 gpd/ft^2
Peak hydraulic loading[a]	700 - 1,200 gpd/ft^2
Average solids loading[a]	14 - 29 lb/d/ft^2
Peak solids loading[a]	30 - 48 lb/d/ft^2
Depth	12 - 15 ft
Weir loading rate	10,000 - 30,000 gpd/linear ft
Maximum upflow velocity in vincinity of weir	12 - 24 ft/hr

[a] In activated sludge systems.

Costs. The cost for rectangular clarifiers are the same as those for circular. Purchased equipment and installation cost for estimation of the total capital investment includes sludge return and waste pumps. Spare pumps (nonclog centrifugal) are also included as necessary. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Service life	40 yr
Overflow rate of flocculator-type clarifier	600 gpd/ft²
Concentration of sludge	1% solids
Pump head	10 ft TDH

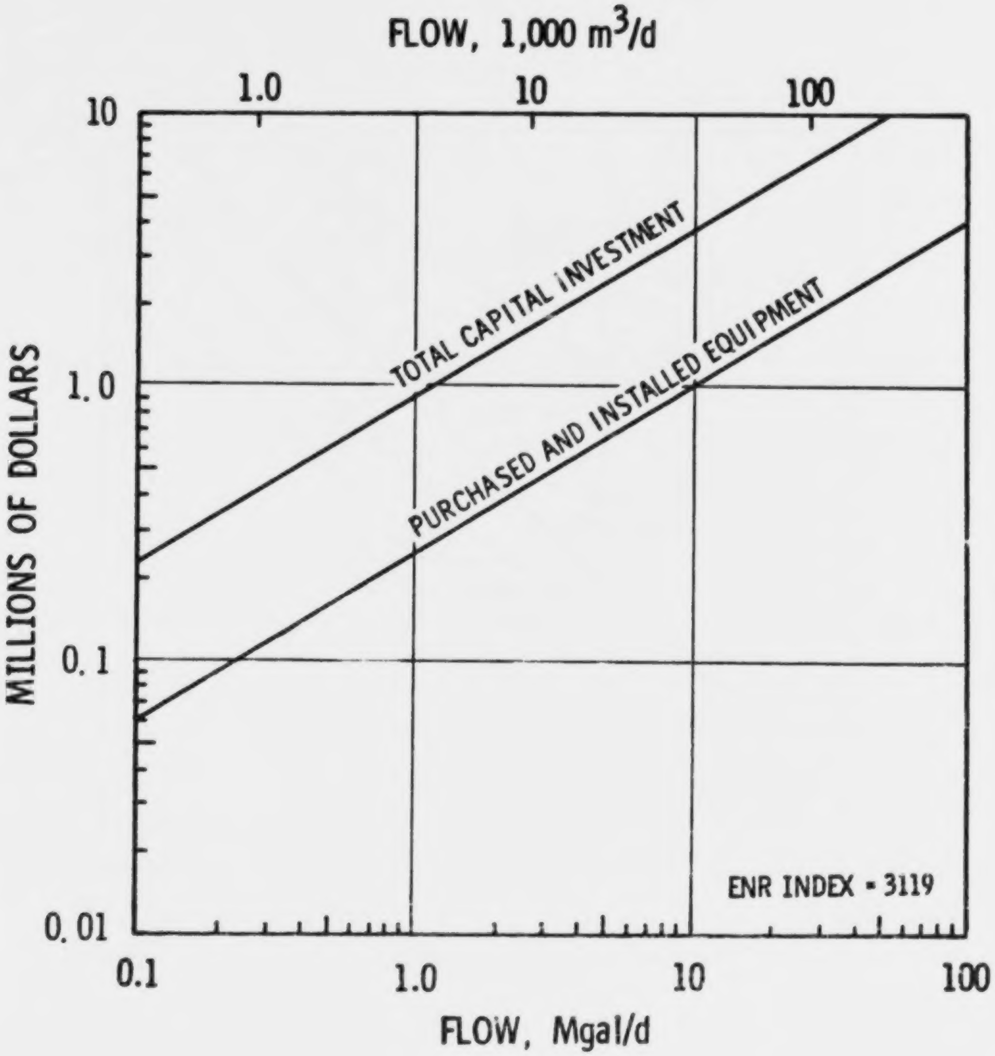
To adjust costs for an alternative overflow rate, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{600 \text{ gpd/ft}^2}{\text{new design overflow rate}}$$

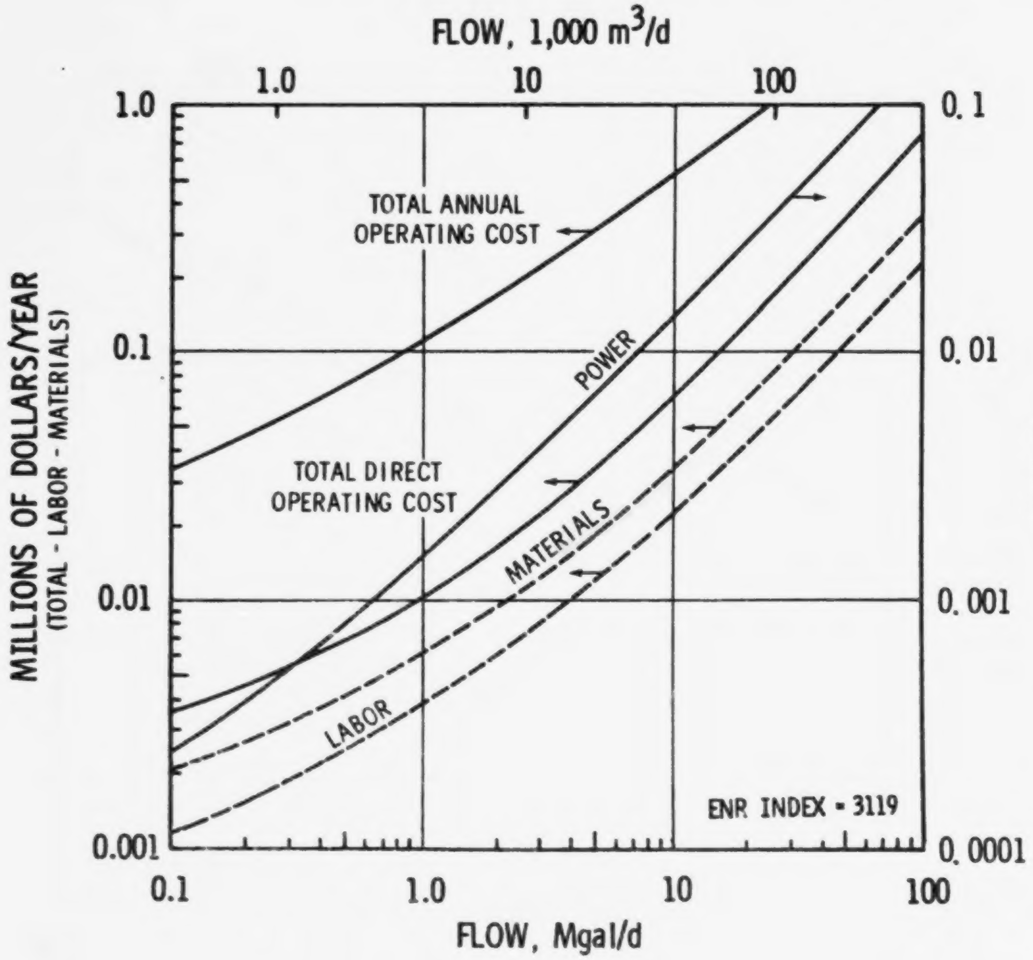
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.3 CLARIFICATION/SEDIMENTATION WITH CHEMICAL ADDITION

Clarification/sedimentation with chemical addition is utilized to remove collodial solids, and for phosphate removal. Sedimentation with chemical addition of alum, ferric chloride, lime, and polymer are discussed in Sections IV.3.3.1 through IV.3.3.4, respectively. The sedimentation process (assumed to be circular primary clarification for cost estimation purposes) was discussed in detail in Section IV.3.2.1.

IV.3.3.1 Alum Addition [1]

Description. Alum or filter alum [$Al_2(SO_4)_3 \bullet 14H_2O$] is a coagulant which, when added to wastewater, initially forms hydrometal complexes which are readily adsorbed by colloids to destabilize them causing flocculation, and finally forms an aluminum hydroxide precipitate. Phosphorus will also be precipitated as the aluminum salt prior to colloid destabilization. The reaction is pH dependent with the minimum solubility occurring at a pH of approximately 6. Since alum acts as an acid, alkalinity will be destroyed after addition to wastewater.

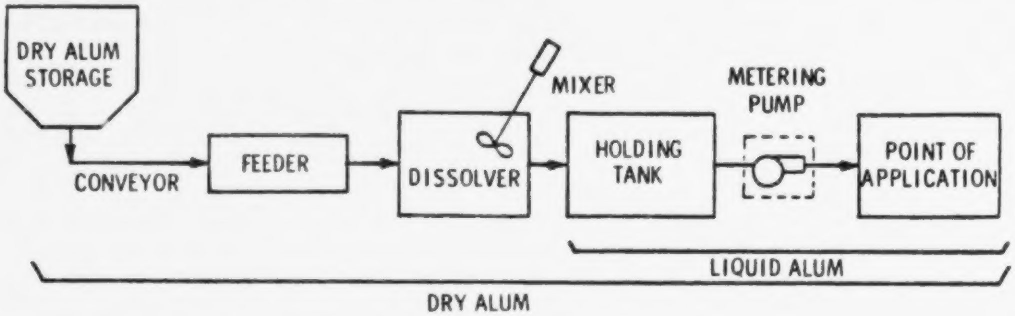
Alum is an off-white crystal which when dissolved in water produces acidic conditions. As a solid, alum may be supplied in lumps, or in ground, rice, or powdered form. Shipments may be in small bags (100 lb), in drums, or in bulk quantities (over 40,000 lb). In liquid form, alum is commonly supplied as a 50 percent solution delivered in minimum loads of 4,000 gallons. The choice between liquid or dry alum use is dependent on factors such as availability of storage space, method of feeding, and economics. In general, purchase of liquid alum is justified only when the supplier is close enough to make differences in transportation costs negligible. Dry alum is stored in mild steel or concrete bins with appropriate dust collection equipment. Because dry alum is slightly hydroscopic, provisions are made to avoid moisture, which could cause caking and corrosive conditions.

Before addition to wastewater, dry alum must be dissolved, forming a concentrated solution. Bulk-stored or hopper-filled alum is transported to a feeder mechanism by bucket elevator, screw conveyor or a pneumatic device. Three basic types of feeders are in common use: volumetric, belt gravimetric, and loss-in-weight gravimetric. The feeder supplies a controlled quantity of dry alum (accuracy ranges from about 1% to 7%) to a mixed dissolver vessel. Because alum solubility is temperature dependent, the quantity supplied depends on the concentrate strength desired and the temperature.

Because alum solution is corrosive, the dissolving chamber as well as the following storage tanks, pumps, piping and surfaces that may come in contact with the solution or generated fumes must be constructed of resistant materials such as type 316 stainless steel, fiberglass reinforced plastic (FRP), or plastics.

Rubber or saran-lined pipes are commonly used. Liquid alum, which crystallizes at about 30°F and freezes at about 18°F, is stored and shipped in insulated type 316 stainless steel or rubber-lined vessels. Feeding of liquid alum (purchased or made up on site) to wastewater treatment unit processes may be accomplished by gravity, pumping, or using a roto-dip feeder. Diaphragm pumps and valves are common.

Equipment normally associated with alum addition includes bins, hoppers, conveyors and elevators, liquid storage tanks, dry and wet feeders, and pH instrumentation. A flow diagram of an alum addition process is shown below.



Typical design criteria. Typical design criteria for primary circular clarification are described in Section IV.3.2.1. Typical design criteria for alum addition are shown below.

Design criteria	Value
Dosage, determined by jar testing	Generally 5 - 20 mg/L as aluminum as an average for most industrial waste
In mixing,	
G[a]	Approximately 300/s
t[b]	≤30 s
In flocculation,	
Gt	Approximately 10^5
GCt	Approximately 100
In sedimentation,	
Overflow rate	500 - 600 gpd/ft^2 average, 800 - 900 gpd/ft^2 (peak)
Alum unit cost	$185/ton

[a] Measure of mean velocity gradient.

[b] Time.

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps, clarifiers, liquid alum, chemical feed equipment sized for twice the average feed rate and storage for at least 15 days, building (except for plants with a capacity less than 1 Mgal/d), rapid mix tank (includes stainless steel mixer), and basins. The rapid mix tank is constructed of concrete, and multiple basins are used for volumes greater than 1500 ft^3. Costs are based on the addition of alum into the primary circular clarification system. An alum dosage of 200 mg/L as $Al_2(SO_4)_3 \cdot 14H_2O$ was assumed for cost estimation.

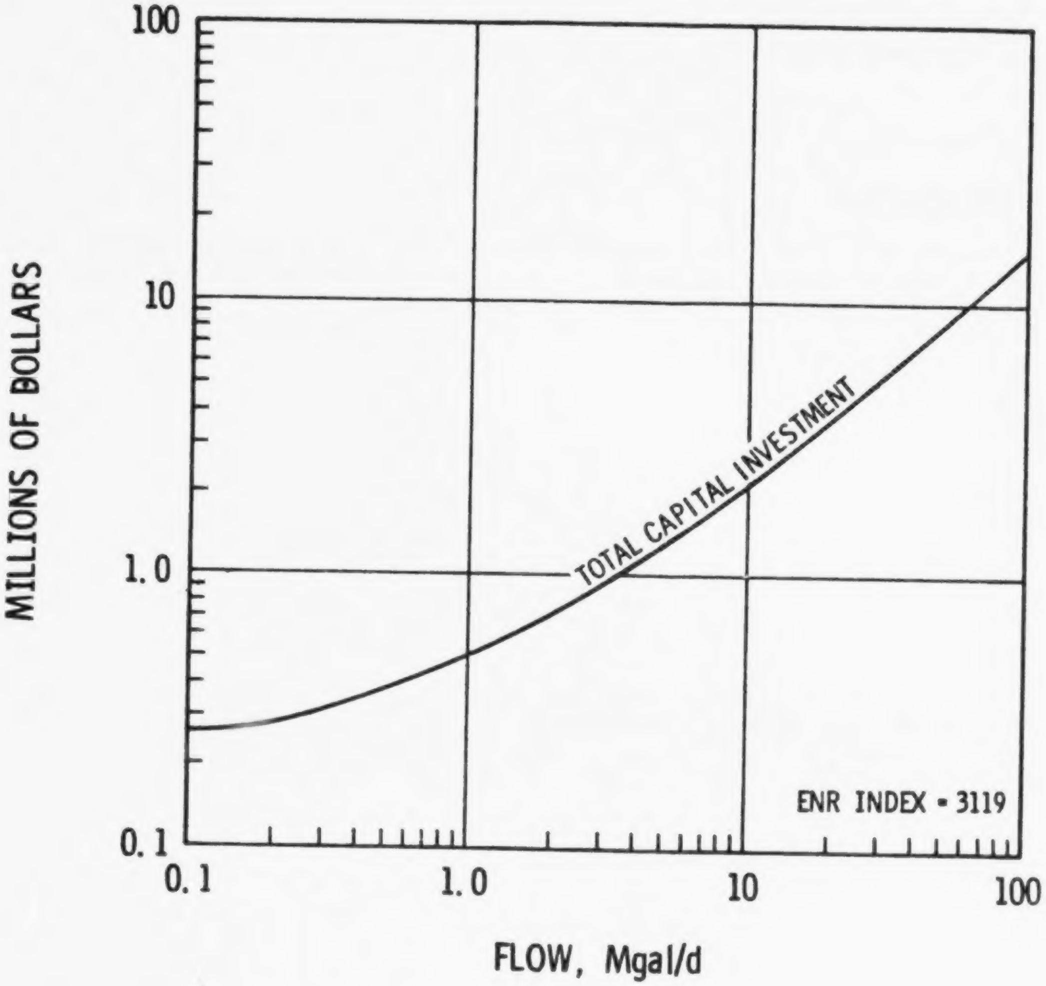
To adjust costs at the effective flow (Q_E) for different alum dosages, use the following equation:

$$Q_E = Q_{Design} \times \frac{\text{Alum dosage}}{200 \text{ mg/L}}$$

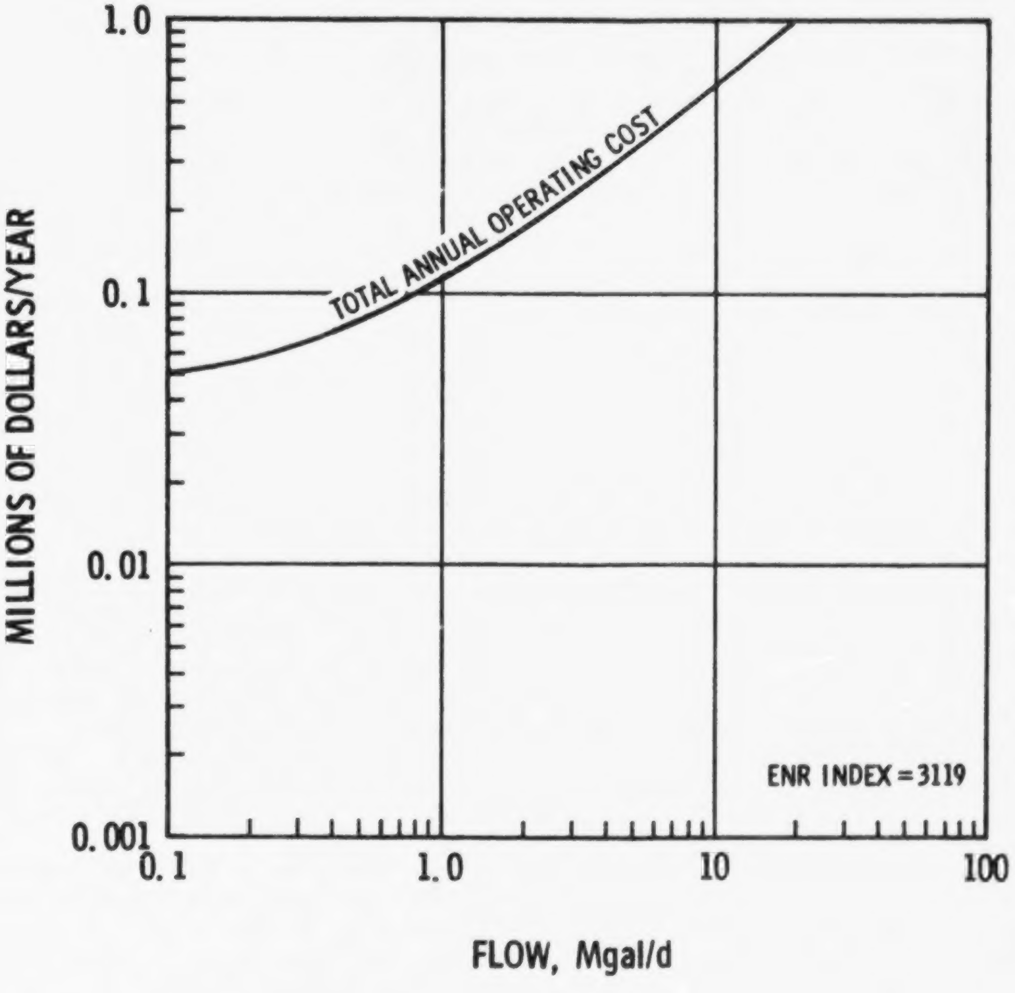
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



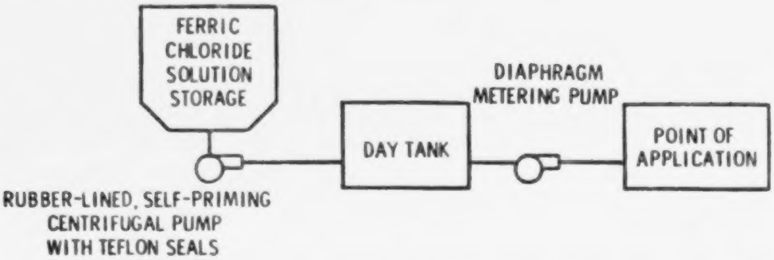
IV.3.3.2 Ferric Chloride Addition [1]

Description. Ferric chloride ($FeCl_3$) is a chemical coagulant which, when added to wastewater destabilizes colloids and reacts with alkalinity and phosphates, similar to alum, forming insoluble iron salts. The colloidal particle size of insoluble ferric phosphate is small, requiring excess dosages of ferric chloride to produce a well flocculated iron hydroxide precipitate, which carries the phosphate precipitate. Large excesses of ferric chloride, and corresponding quantities of alkalinity, are required to assure phosphate removal. Exact ferric chloride dosages are usually best determined using jar tests and full-scale evaluations.

Ferric chloride is available in either dry (hydrated or anhydrous) or liquid form. Liquid ferric chloride is a dark brown oil-appearing solution supplied in concentrations ranging between 35 and 45 percent ferric chloride. Because higher concentrations of ferric chloride have higher freezing points, lower concentrations are supplied during winter. Liquid ferric chloride is shipped in 3,000- to 4,000-gallon bulk truckload lots, in 4,000- to 10,000-gallon carloads, and in 5- to 13-gallon carboys. Ferric chloride solution stains surfaces which it contacts and is highly corrosive (a one percent solution has a pH of 2.0); consequently, it must be stored and handled with care. Storage tanks are equipped with vents and vacuum relief valves. Tanks are constructed of fiberglass reinforced plastic, rubber-lined steel and plastic-lined steel. Because of freezing potential, ferric chloride solutions are either stored in heated areas or in heated and insulated vessels in northern climates.

Ferric chloride solution should not be diluted because of possible unwanted hydrolysis. Consequently, feeding at the concentration of the delivered product is common. The stored solution is transferred to a day tank using graphite or rubber-lined self-priming centrifugal pumps with corrosion resistant Teflon seals. From the day tank, controlled quantities are fed to the unit process using Rotodip feeders or diaphragm metering pumps. Rotometers are not used for ferric chloride flow measurement because the material tends to deposit on and stain the glass tubes. All pipes, valves, or surfaces that come in contact with ferric chloride must be made of corrosion-resistant materials such as rubber or Saran lining, Teflon, or vinyl. Similar treatment results are obtainable by substituting ferrous chloride, ferric sulfate, ferrous sulfate, or spent pickle liquor for ferric chloride. Details of storage feeding and control for these materials are similar to those for ferric chloride. Dry ferric chloride may also be dissolved on site before use in treatment.

Equipment normally associated with the ferric chloride addition process includes liquid storage tanks, dry and wet feeders, and pH instrumentation. A flow diagram of a ferric chloride addition process is shown below.


FERRIC CHLORIDE SOLUTION STORAGE
DIAPHRAGM METERING PUMP
DAY TANK
POINT OF APPLICATION
RUBBER-LINED, SELF-PRIMING CENTRIFUGAL PUMP WITH TEFLON SEALS

Typical design criteria. Typical design criteria for primary circular clarification are described in Section IV.3.2.1.2. Typical design criteria for ferric chloride addition are shown below.

Design criterion	Value
Dosage, determined by jar testing	20 - 100 mg $FeCl_3$/L are common for most industrial wastes
In mixing,	
G	Approximately 300/s
t	≤30 s
$FeCl_3$ unit cost	\$38/ton

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps, clarifiers, liquid ferric chloride, chemical feed equipment sized for twice the average feed rate, storage for at least 15 days, rapid mix tank, stainless steel mixer, and building (except for plants with a capacity less than 1 Mgal/d). The rapid mix tank is constructed of concrete, and multiple basins are used for volumes greater than 1,500 ft³. Costs are based on primary circular clarification with ferric chloride addition. A ferric chloride dosage of 100 mg/L was assumed for cost estimation.

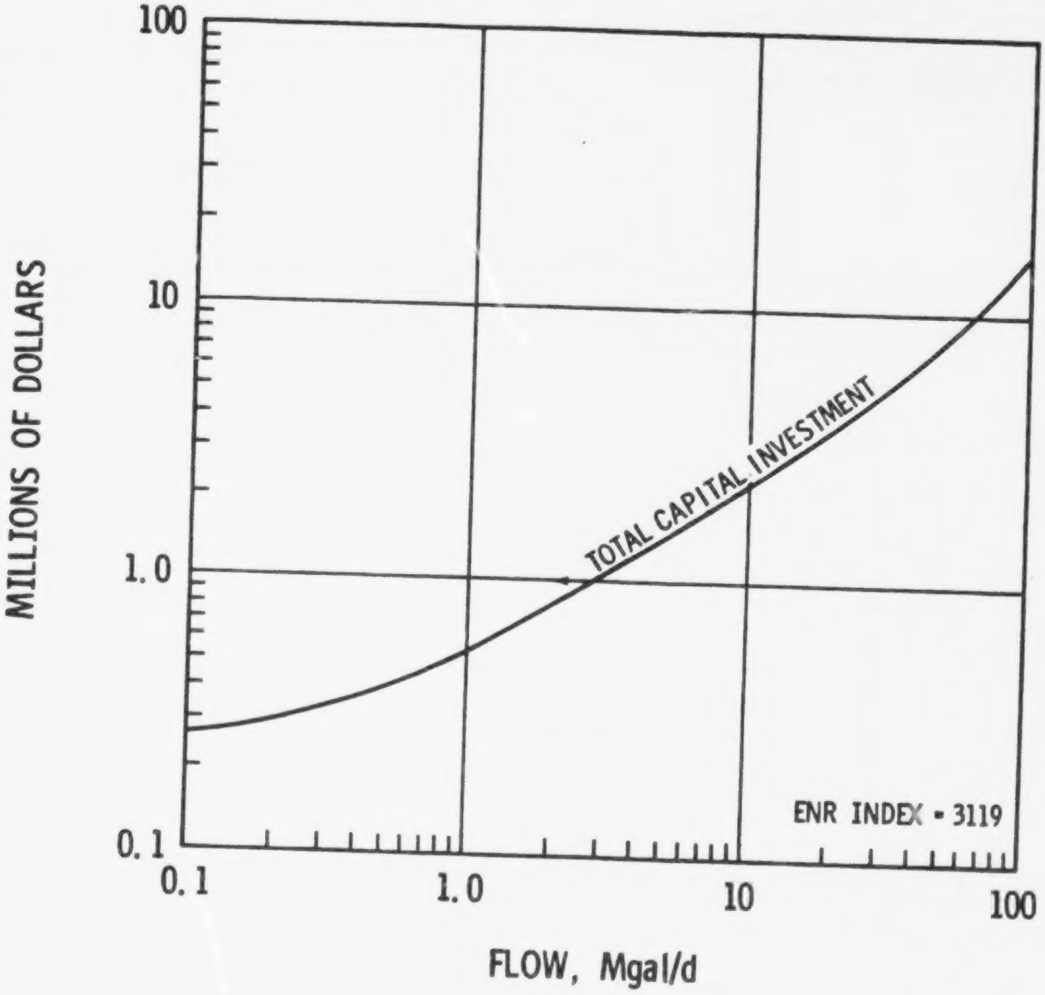
To adjust costs at the effective flow (Q_E) for different $FeCl_3$ dosages, use the following equation:

$$Q_E = Q_{Design} \times \frac{FeCl_3 \text{ dosage}}{100 \text{ mg/L}}$$

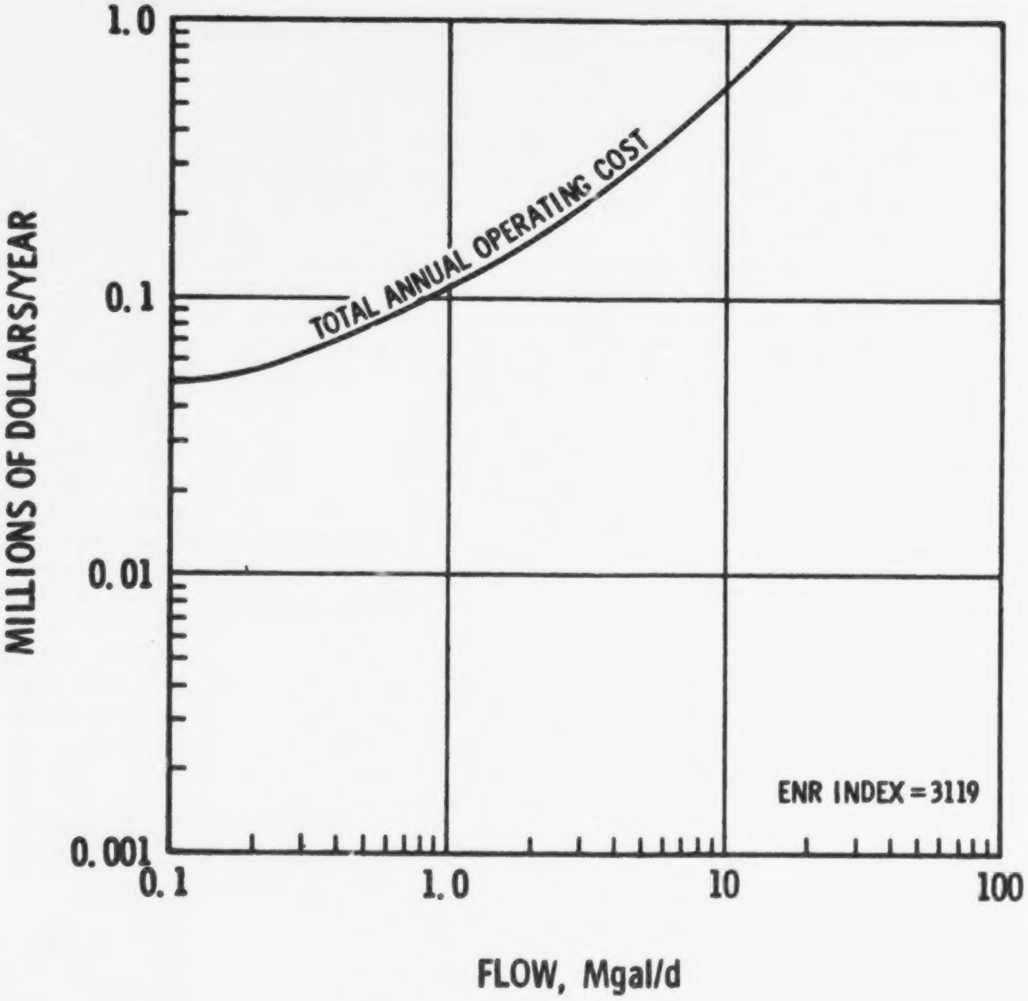
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



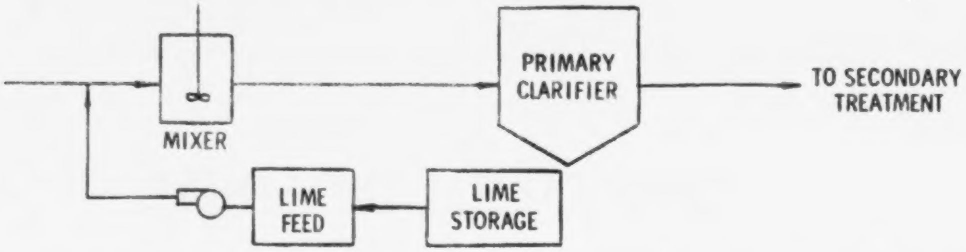
IV.3.3.3 Lime Addition (Primary) [1]

Description. Lime clarification of raw wastewater removes suspended solids as well as phosphates. There are two basic processes: the low-lime system and the high-lime system. The low-lime process consists of the addition of lime to obtain a pH of approximately 9 to 10. Generally, a subsequent biological treatment system is capable of readjusting the pH through natural recarbonation. The high-lime process consists of the addition of lime to obtain a pH of approximately 11 or more. In this case, the pH generally requires readjusting with carbon dioxide or acid to be acceptable to the secondary treatment system.

Lime can be purchased in many forms; quicklime (CaO) and hydrated lime [$Ca(OH)_2$] are the most prevalent forms. In either case, lime is usually purchased in the dry state, in bags, or in bulk. Bulk lime can be (1) shipped by trucks that are generally equipped with pneumatic unloading equipment; or (2) shipped by rail cars that consist of covered hoppers. The rail cars are emptied by opening a discharge gate, which discharges to a screw conveyor. The bulk lime is then transferred by the screw conveyor to a bucket elevator, which empties into the elevated storage tank. Bulk storage usually consists of steel or concrete bins. Storage vessels should be water- and air-tight to prevent the lime from "slaking".

Lime is generally made into a wet suspension or slurry before being introduced into the treatment system. The precise steps involved in converting from the dry to the wet stage will vary according to the size of operation and type and form of lime used. In the smallest plants, bagged hydrated lime is often charged manually into a batch mixing tank with the resulting "milk-of-lime" (or slurry) being fed by means of a so-called solution feeder to the process. Where bulk hydrate is used, some type of dry feeder charges the lime continuously to either a batch or continuous mixer, then, by means of solution feeder, to the point of application. With bulk quicklime, a dry feeder is also used to feed a slaking device, where the oxides are converted to hydroxides, producing a paste or slurry. The slurry is then further diluted to milk-of-lime before being piped by gravity or pumped to the process. Dry feeders can be of the volumetric or gravimetric type.

Equipment normally associated with the lime clarification process includes bins, hoppers, conveyors and elevators, liquid storage tanks, dry and wet feeders, lime slakers, and pH instrumentation. A flow diagram of a lime clarification process is shown on the next page.



Typical design criteria. Typical design criteria for primary circular clarification are described in Section IV.3.2.1.2. The lime requirements of a lime clarification system are tabulated below.

Feed water alkalinity, mg/L (as $CaCO_3$)	Clarifier pH	Approximate lime dose, mg/L (as CaO)
300	9.5	185
300	10.5	270
400	9.5	230
400	10.5	380

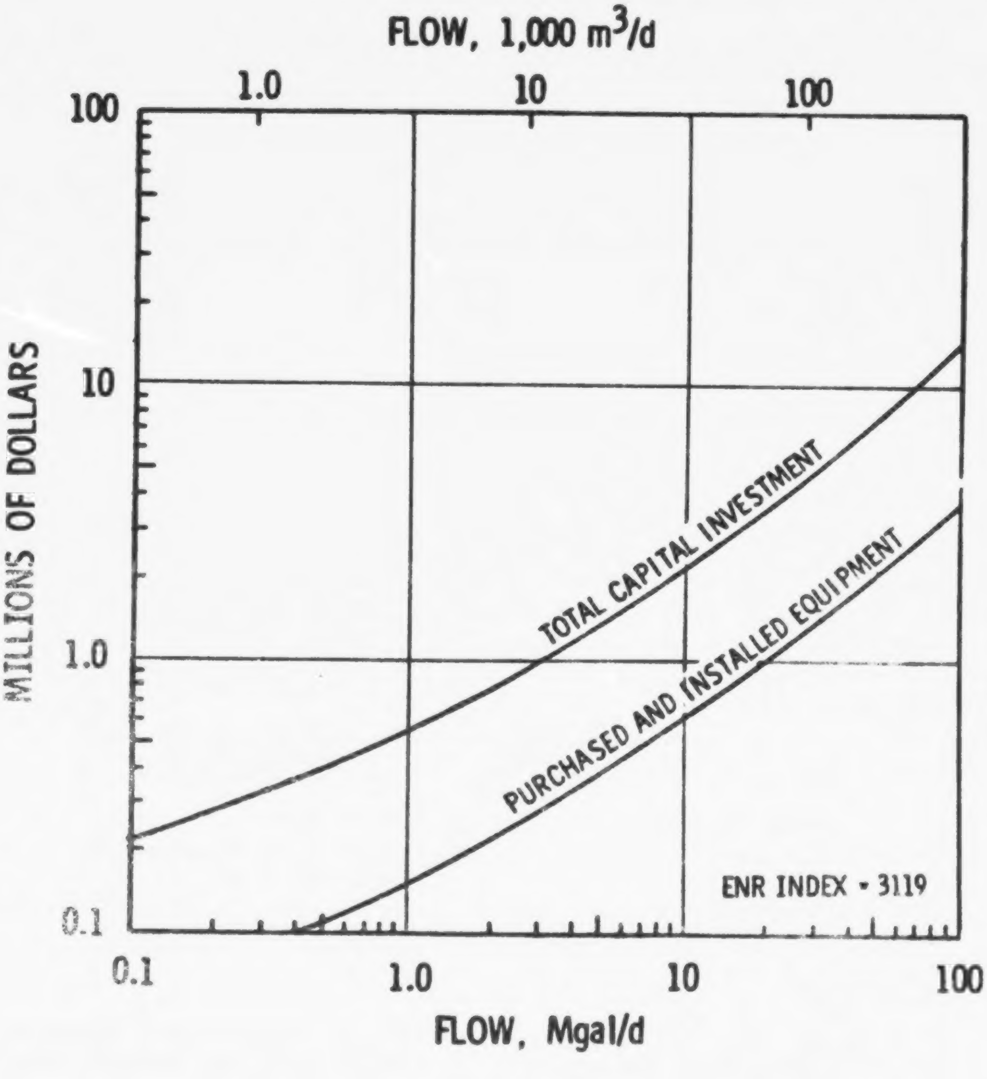
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps, clarifiers, chemical storage and feeding equipment, and hydrated lime for 0.1 to 10 Mgal/d plants or pebble quicklime for 10 to 100 Mgal/d plants. Piping and buildings to house the feeding equipment are not included. Costs are based on primary circular clarification with lime addition. The following lime addition operating characteristics were additionally assumed for cost estimation:

Operating characteristic	Assumed value
Lime dosage	150 mg/L
Storage	150 days minimum
CaO	$54.00/ton

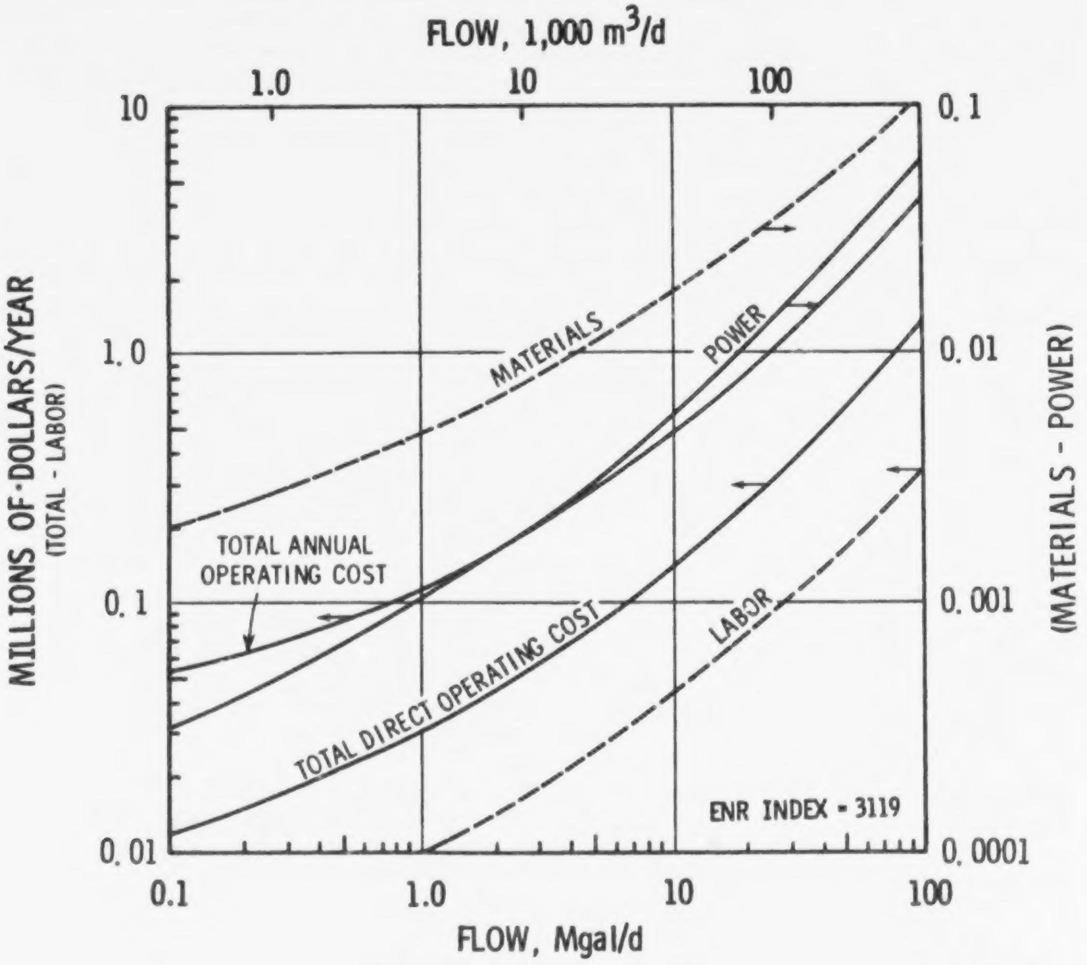
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



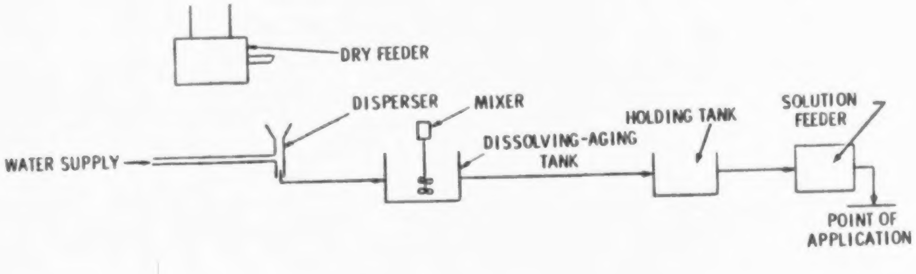
IV.3.3.4 Polymer Addition [1]

Description. Polymers or polyelectrolytes are high-molecular-weight compounds (usually synthetic) which, when added to wastewater, can be used as coagulants, coagulant aids, filter aids, or sludge conditioners. In solution, polymers may carry either a positive, negative, or neutral charge and, as such, they are characterized as cationic, anionic, or nonionic. As a coagulant or coagulant aid, polymers act as bridges, reducing charge repulsion between colloidal and dispersed floc particles, and increasing settling velocities. As a filter aid, polymers strengthen fragile floc particles, controlling filter penetration and reducing particle breakthrough. Filterability and dewatering characteristics of sludges may similarly be improved through the use of polyelectrolytes.

Polymers are available in predissolved liquid or dry form. Dry polymers are supplied in relatively small quantities (up to about 100-lb bags or barrels) and must be dissolved on site prior to use. A stock solution, usually about 0.2 to 2.0 percent concentration, is made up for subsequent feeding to the treatment process. Preparation involves automatic or batch wetting, mixing, and aging. Stock polymer solutions may be very viscous.

Surfaces coming in contact with the polymer stock solution should be constructed of resistant materials such as type 316 stainless steel, fiberglass-reinforced plastic or other plastic lining materials. Polymers may be supplied as a prepared stock solution ready for feeding to the treatment process. Many competing polymer formulations with different characteristics are available, requiring somewhat differing handling procedures. Manufacturers should be consulted for optimum practices. Polymer stock solutions are generally fed to unit processes using equipment similar to that commonly in service for dissolved coagulant addition. Because of the high viscosity of stock solutions, special attention should be paid to the diameter and slopes of pipes, as well as the size of orifices used in the feed systems.

Equipment normally associated with the polymer addition process includes bins, hoppers, liquid storage tanks, and dry and wet feeders. A flow diagram of a polymer addition process is shown below.



Typical design criteria. Typical design criteria for primary circular clarification are described in Section IV.3.2.1.2. Typical design criteria for polymer addition are shown below.

Design criteria	Value
Dosage	Determined by jar testing
Materials contacting polymer solutions	Type 316 stainless steel, FRP or plastic
Storage conditions	Cool and dry; storage period should be minimized
Viscosity	Must be considered in feeding system design

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps; clarifier; chemical storage, chemical feeding, and rapid mixing tanks; piping and building to house the feeding equipment; and bag storage. Cost varies according to the plant size as shown below.

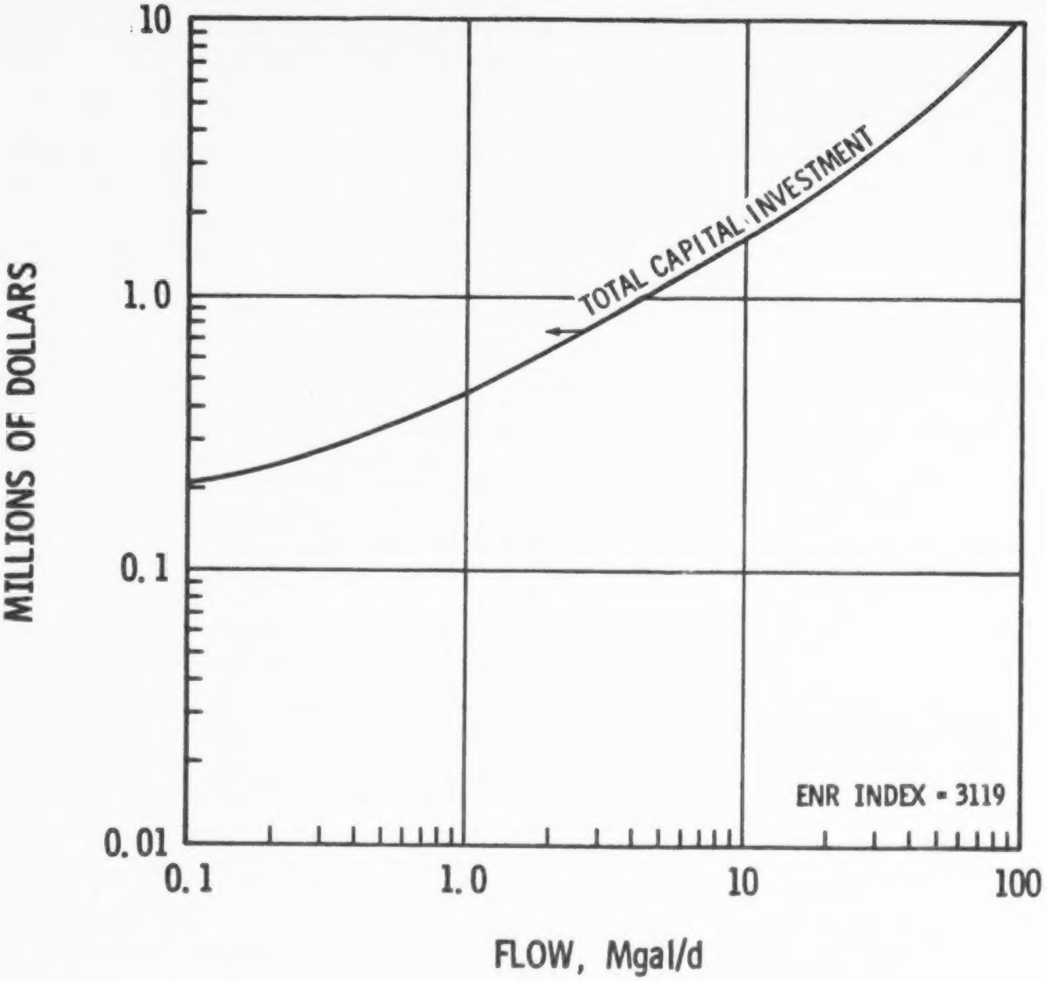
Plant size	Factors affecting costs
0.1 Mgpd	Manual operation of feeder, mix tank, solution feeder, and holding tank are included in cost estimation; no separate building is required.
1 Mgpd and smaller	Manual procedures are applied; two systems of tanks and feeders are included in cost estimation.
10 Mgpd	Cost estimate includes feeders and mixing tanks, one day tank, and two solution feeders.
100 Mgpd	Cost estimate includes four feeders and mixing tanks, two holding tanks, and 10 solution feeders; rapid mix tank is concrete and equipped with stainless steel mixer and handrails.

This cost assumption is based on primary circular clarification with polymer addition at a dosage of 1.0 mg/L and a cost of $1.50/lb.

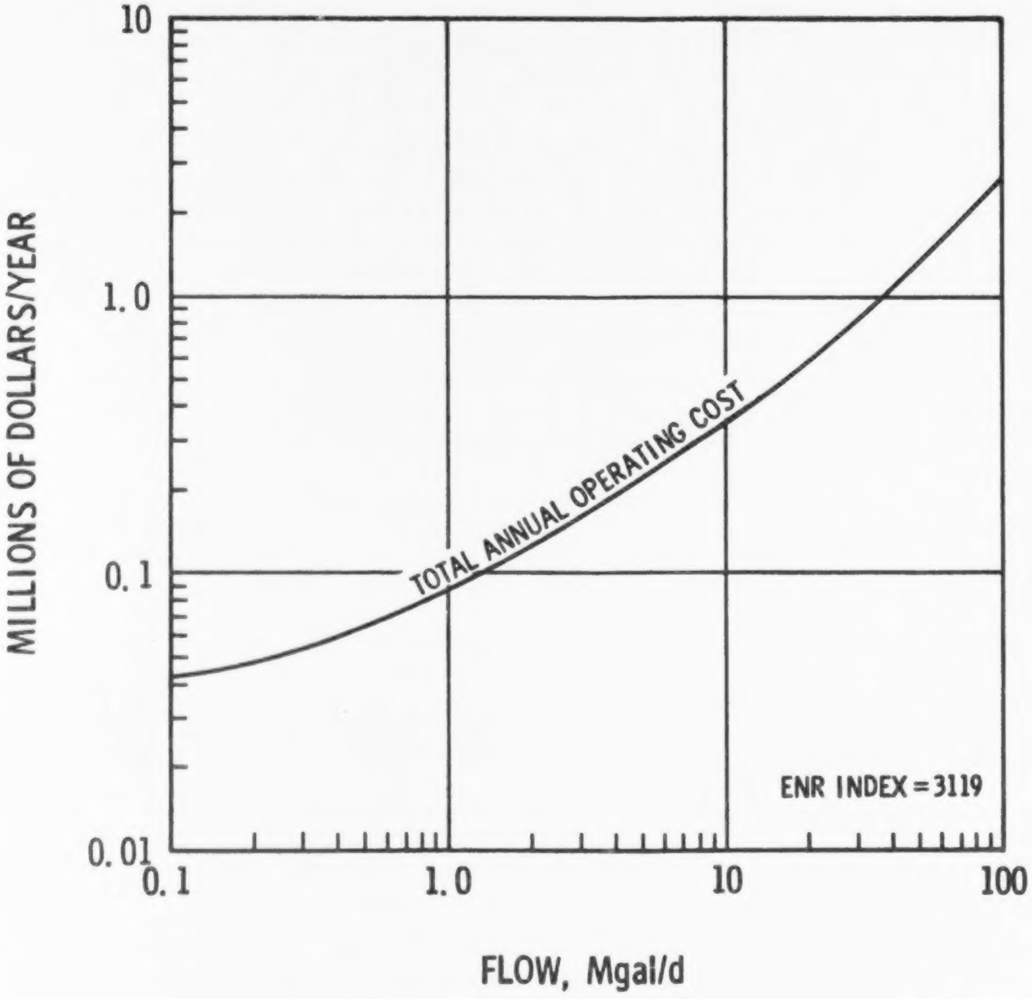
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.4 DISSOLVED AIR FLOTATION (GAS FLOTATION)

IV.3.4.1 Description

Dissolved air flotation (DAF) is used to remove suspended solids by using highly pressurized air to form bubbles. DAF consists of saturating a portion or all of the wastewater feed, or a portion of recycled effluent, with air at a pressure of 25 to 70 lb/in^2 (gage). The pressurized wastewater is held at this pressure for 0.5 to 3.0 minutes in a retention tank and then released to the flotation chamber. The sudden reduction in pressure results in the release of microscopic air bubbles in the flotation chamber, which attach themselves to oil and suspended particles in the wastewater. This results in agglomeration which, due to the entrained air, results in greatly increased vertical rise rates of about 0.5 to 2.0 ft/min. The floated materials rise to the surface to form a froth layer. Specially designed flight scrapers or other skimming devices continuously remove the froth. The retention time in the flotation chamber is usually about 20 to 60 minutes. The effectiveness of dissolved air flotation depends on the attachment of bubbles to the suspended oil and other particles that are to be removed from the waste stream. The attraction between the air bubble and particle is primarily a result of the particle surface charges and bubble-size distribution.

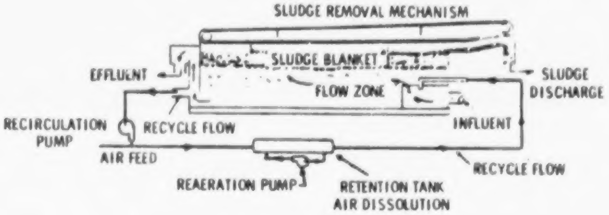
The more uniform the distribution of water and microbubbles, the shallower the flotation unit can be. Generally, the depth of effective flotation units is between 4 and 9 feet.

In certain cases, the surface sludge layer can attain a thickness of many inches and can be relatively stable for a short period. The layer thickens with time, but undue delays in removal will cause a release of particulates back to the liquid.

DAF units can be round, square, or rectangular. In addition, gases other than air can be used. The petroleum industry has used nitrogen, with closed vessels, to reduce the possibilities of fire. Chemical addition is normally used for coagulation colloidal solids and for breaking emulsions.

Common modifications include chemical additions such as alum, ferric chloride, lime and polymer. The chemicals can be added to aid in the coagulation process prior to the actual flotation step. Capital cost will be greater for gas flotation with chemical addition than for gas flotation alone. This is due to additional equipment and installation costs. Such equipment will include chemical feed equipment, rapid mix tank, chemical storage and a stainless steel mixer.

Equipment normally associated with gas flotation system includes dissolved air flotation units, air compressors, and skimmers. A flow diagram of a gas flotation system is shown below.


SLUDGE REMOVAL MECHANISM
SLUDGE BLANKET
EFFLUENT
FLOW ZONE
SLUDGE DISCHARGE
RECIRCULATION PUMP
RECYCLE FLOW
INFLUENT
AIR FEED
RECYCLE FLOW
REAERATION PUMP
RETENTION TANK AIR DISSOLUTION

(SEE SECTION IV.7)

IV.3.4.2 Typical design criteria

Typical design criteria for gas flotation systems are as follows:

Design criteria	Values
Pressure	25 - 70 lb/in.^{2}[a]
Air-to-solids ratio	0.01 - 0.1 lb/lb
Float detention	20 - 60 min
Surface hydraulic loading	500 - 8,000 gpd/ft^2
Recycle (where employed)	5 - 120%

[a] Pressure in psig.

IV.3.4.3 Costs

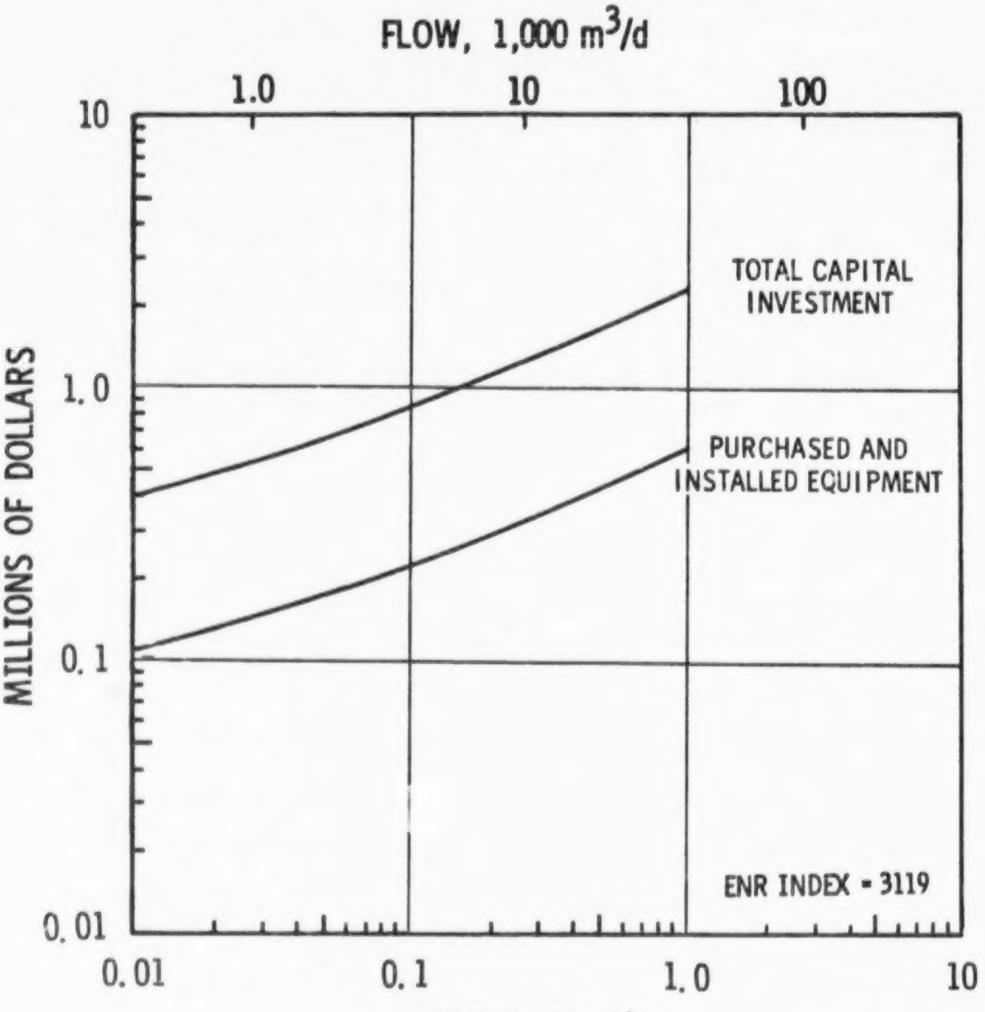
Purchased equipment and installation cost for estimation of the total capital investment includes the DAF unit and equipment to feed in coagulants if needed. The following design characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Air injection	1.25 ft^3/1,000 gal
Recycle	33%
Float detention time	25 min
Surface hydraulic loading	500 - 8,000 gpd/ft^2

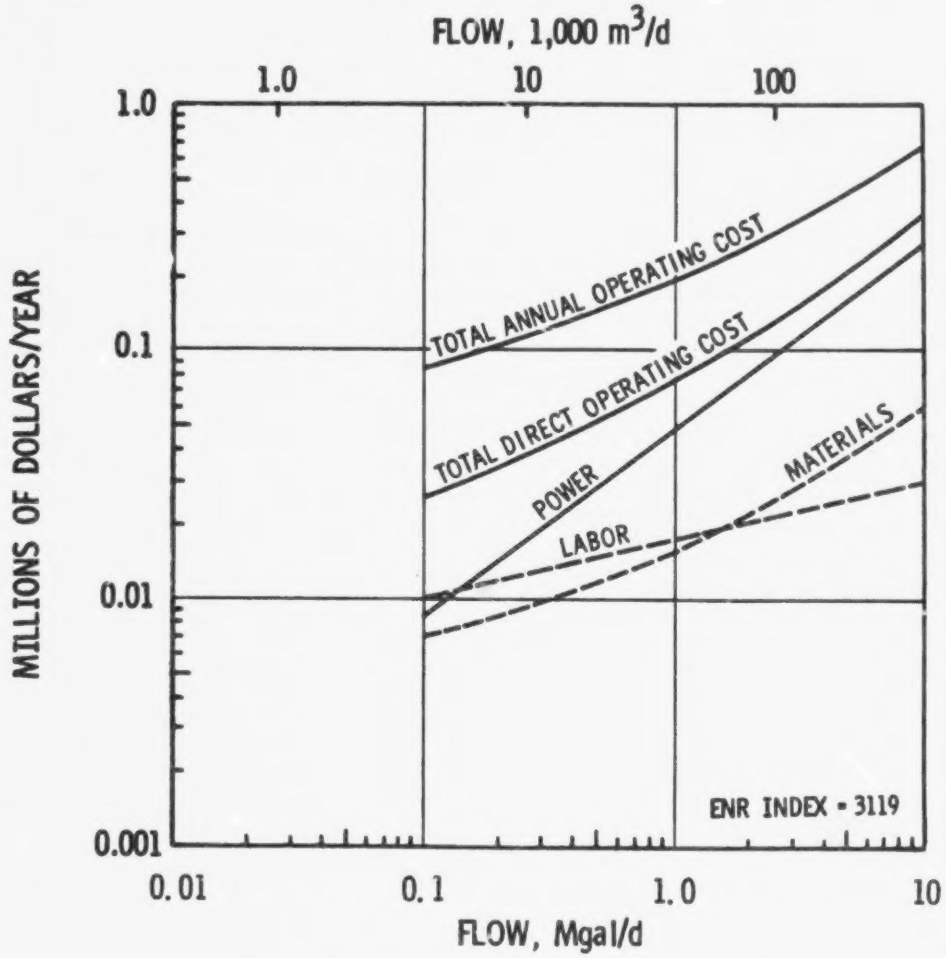
IV.3.4.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.5 DISSOLVED AIR FLOTATION WITH CHEMICAL ADDITION

Gas flotation is used to remove lighter suspended materials whose specific gravity is only slightly in excess of 1.0. It is usually used to remove oil and grease materials; the process is also used when existing clarifiers are overloaded hydraulically because of its lower surface area requirement. Alum ($Al_2(SO_4)_3 \bullet 14\ H_2O$), ferric chloride ($FeCl_3$), and polymers can be added to aid in the coagulation process prior to the actual flotation step. The basic gas flotation system is described in Section IV.3.4. The typical design criteria and descriptions of chemical addition (alum, ferric chloride, lime and polymer) are presented in Section IV.3.3.1 through IV.3.3.4. Costs are estimated for the combination of gas flotation with chemical addition. The exact chemical costs are broken down in Section IV.3.4 and dosages indicated.

IV.3.5.1 Alum Addition in Gas Flotation [1]

Description. Alum or filter alum [$Al_2(SO_4)_3 \bullet 14H_2O$] is a coagulant which, when added to wastewater, reacts with available alkalinity (carbonate, bicarbonate and hydroxide) and phosphate to form insoluble aluminum salts. The combination of alum with alkalinity or phosphate are competing reactions that are pH dependent.

Alum is an off-white crystal which when dissolved in water produces acidic conditions. As a solid, alum may be supplied in lumps, or in ground, rice, or powdered form. Shipments may be in small bags (100 lb), in drums or in bulk quantities (over 40,000 lb). In liquid form, alum is commonly supplied as a 50 percent solution delivered in minimum loads of 4,000 gallons. The choice between liquid or dry alum use is dependent on factors such as availability of storage space, method of feeding, and economics. In general, purchase of liquid alum is justified only when the supplier is close enough to make differences in transportation costs negligible. Dry alum is stored in mild steel or concrete bins with appropriate dust collection equipment. Because dry alum is slightly hydroscopic, provisions are made to avoid moisture, which could cause caking and corrosive conditions.

Before addition to wastewater, dry alum must be dissolved, forming a concentrated solution. Bulk-stored or hopper-filled alum is transported to a feeder mechansim by bucket elevator, screw conveyor, or a pneumatic device. Three basic types of feeders are in common use: volumetric, belt gravimetric, and loss-in-weight gravimetric. The feeder supplies a controlled quantity of dry alum (accuracy ranges from about 1% to 7%) to a mixed dissolver vessel. Because alum solubility is temperature-dependent, the quantity supplied depends on the concentrate strength desired and the temperature.

Because alum solution is corrosive, the dissolving chamber as well as the following storage tanks, pumps, piping, and surfaces that may come in contact with the solution or generated fumes must be constructed of resistant materials such as type 316 stainless steel, fiberglass reinforced plastic (FRP), or plastic. Rubber or saran-lined pipes are commonly used. Liquid alum, which crystallizes at about 30°F and freezes at about 18°F, is stored and shipped in insulated type 316 stainless steel or rubber-lined vessels. Feeding of liquid alum (purchased or made up on site) to wastewater treatment unit processes may be accomplished by gravity, pumping, or using roto-dip type feeder. Diaphragm pumps and valves are common.

Equipment normally associated with alum addition includes bins, hoppers, conveyors and elevators, liquid storage tanks, dry and wet feeders, and pH instrumentation.

Typical design criteria. Typical design criteria for dissolved air flotation are discussed in Section IV.4.3.2. Typical design criteria for alum addition are shown below.

Design criteria	Value
Dosage, determined by jar testing	Generally 5 - 20 mg/L as aluminum
In mixing,	
G^a	Approximately 300/s
t^b	≤30 s
In flocculation,	
GCt	Approximately 100
Gt	Approximately 10^5

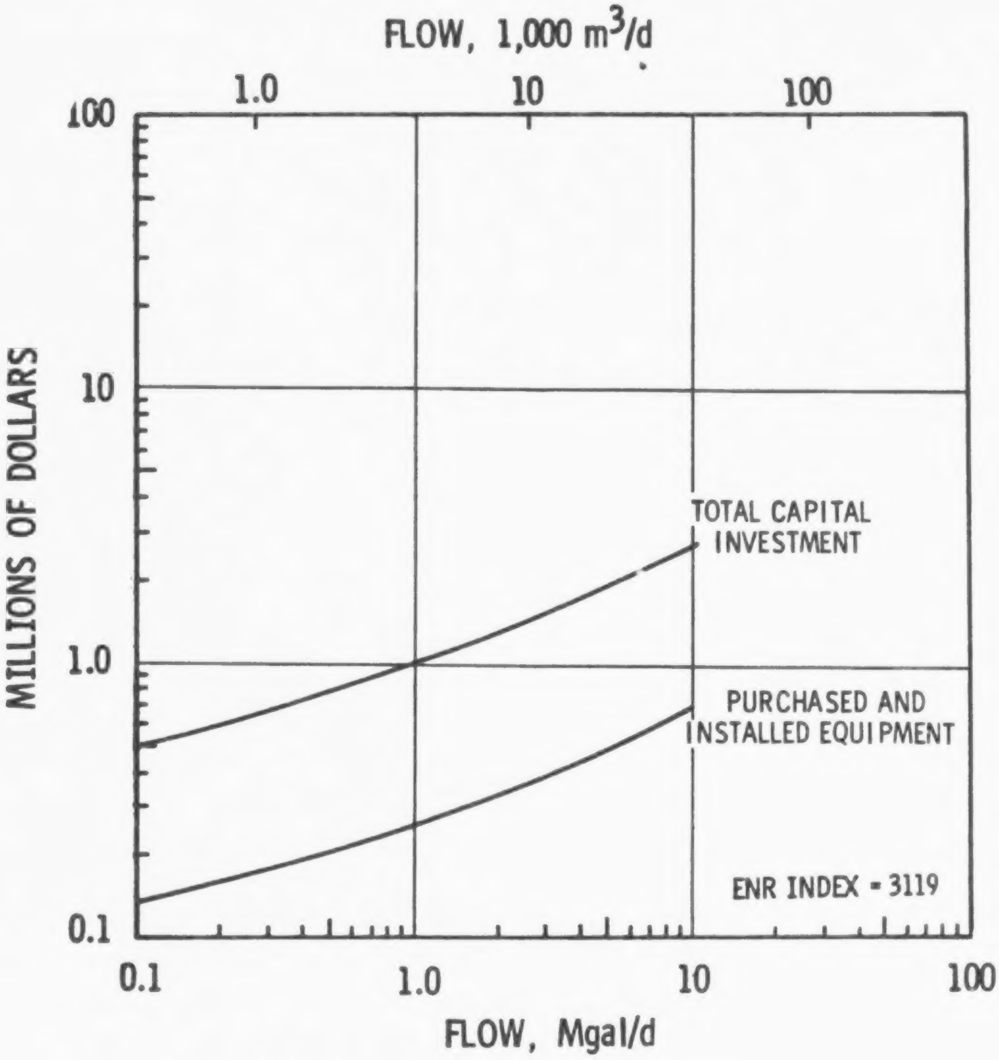
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes the DAF unit, liquid alum (8.3% Al_2O_3), chemical feed equipment sized for twice the average feed rate, storage of at least 15 days, building (except for plants with a capacity less than 1 Mgpd), rapid mix tank, stainless steel mixer, and multiple basins with volumes greater than 1,500 ft³. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Alum dosage	200 mg/L
Air injection	1.25 ft³/1,000 gal
Recycle	33%
Detention time	25 min
Area flow rate	2 - 3 gpm/ft²

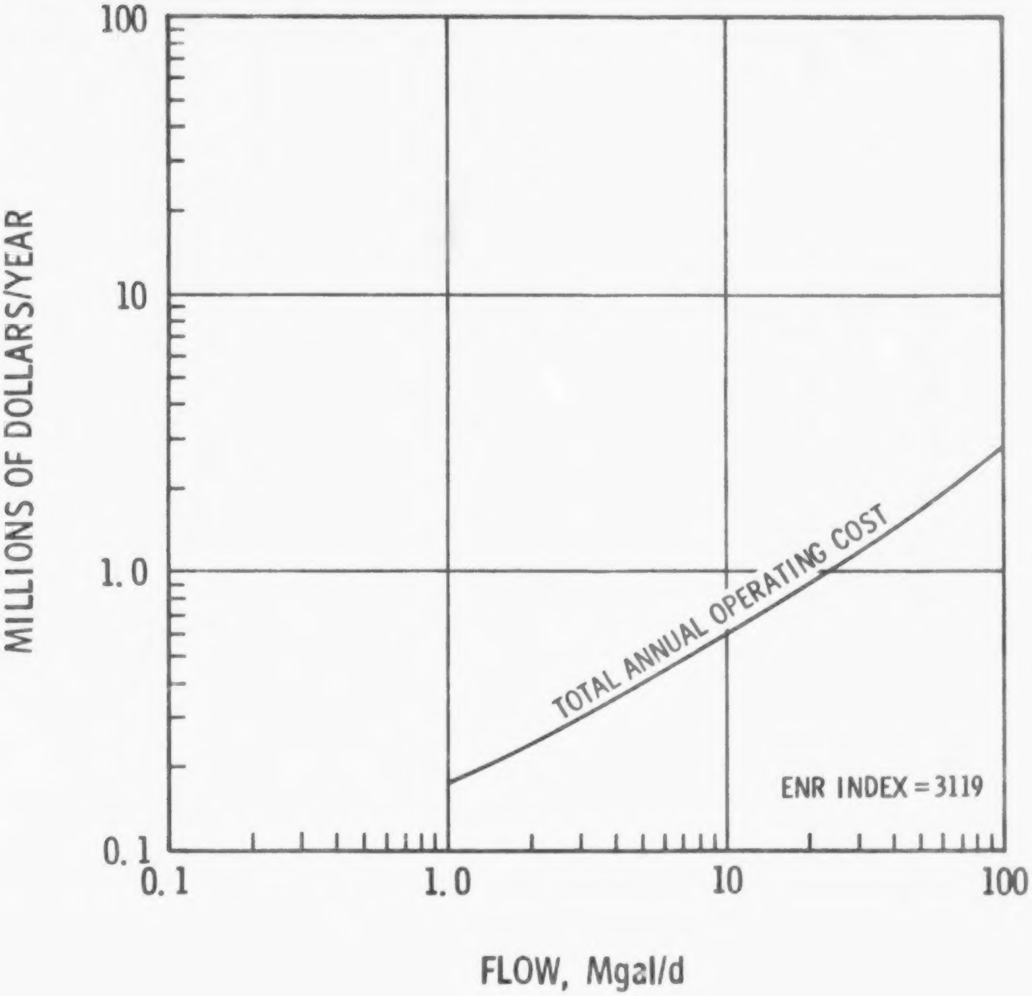
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.5.2 Ferric Chloride Addition in Gas Flotation System [1]

Description. Ferric chloride ($FeCl_3$) is a chemical coagulant which, when added to wastewater, reacts with alkalinity and phosphates, forming insoluble iron salts. The colloidal particle size of insoluble ferric phosphate is small, requiring excess dosages of ferric chloride to produce a well flocculated iron hydroxide precipitate, which carries the phosphate precipitate. Large excesses of ferric chloride, and corresponding quantities of alkalinity, are required to assure phosphate removal. Exact ferric chloride dosages are usually best determined using jar tests and full-scale evaluations.

Ferric chloride is available in either dry (hydrated or anhydrous) or liquid form. Liquid ferric chloride is a dark borwn oil-appearing solution supplied in concentrations ranging between 35 and 45 percent ferric chloride. Because higher concentrations of ferric chloride have higher freezing points, lower concentrations are supplied during winter. Liquid ferric chloride is shipped in 3,000- to 4,000-gallon bulk truckload lots, in 4,000- to 10,000-gallon carloads, and in 5- to 13-gallon carboys. Ferric chloride solution stains surfaces which it contacts and is highly corrosive (a one percent solution has a pH of 2.0); consequently, it must be stored and handled with care. Storage tanks are equipped with vents and vacuum relief valves. Tanks are constructed of fiberglass reinforced plastic, rubber-lined steel and plastic-lined steel. Because of freezing potential, ferric chloride solutions are either stored in heated areas or in heated and insulated vessels in northern climates.

Ferric chloride solution should not be diluted because of possible unwanted hydrolysis. Consequently, feeding at the concentration of the delivered product is common. The stored solution is transferred to a day tank using graphite or rubber-lined self-priming centrifugal pumps with corrosion resistant Teflon seals. From the day tank, controlled quantities are fed to the unit process using Rotodip feeders or diaphragm metering pumps. Rotometers are not used for ferric chloride flow measurement because the material tends to deposit on and stain the glass tubes. All pipes, valves, or surfaces that come in contact with ferric chloride must be made of corrosion-resistant materials such as rubber or Saran lining, Teflon, or vinyl. Similar treatment results are obtainable by substituting ferrous chloride, ferric sulfate, ferrous sulfate, or spent pickle liquor for ferric chloride. Details of storage feeding and control for these materials are similar to those for ferric chloride. Dry ferric chloride may also be dissolved on site before use in treatment.

Equipment normally associated with the ferric chloride addition process includes liquid storage tanks, dry and wet feeders, and pH instrumentation.

Typical design criteria. Typical design criteria for dissolved air flotation are discussed in Section IV.4.3.2. Typical design criteria for ferric chloride addition is shown below.

Design criterion	Value
Dosage, determined by jar testing	20 - 100 mg $FeCl_3$/L are common
In mixing,	
G^a	Approximately 300/s
t^b	≤30 s

[a] Measure of mean velocity gradient.

[b] Time.

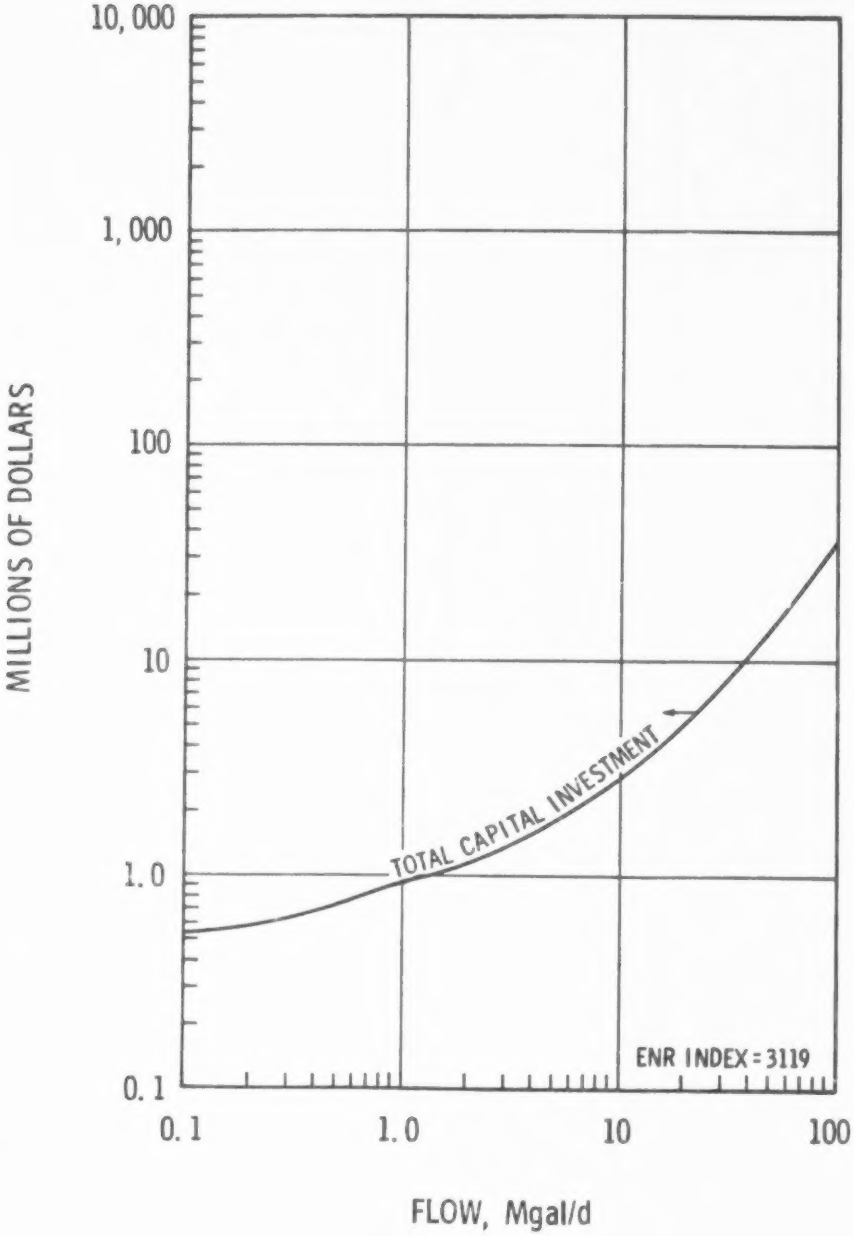
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes the DAF unit, liquid ferric chloride, chemical feed equipment sized for twice the average feed rate, storage of at least 15 days, building (except for plants with a capacity less than 1 Mgpd), rapid mix tanks, multiple basins with volumes greater than 1,500 ft^3, and stainless steel mixer. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
$FeCl_3$	100 mg/L
Air injection	1.25 ft^3/100 gal
Recycle	33%
Detention time	25 min
Area flow rate	2.3 gpm/ft^3

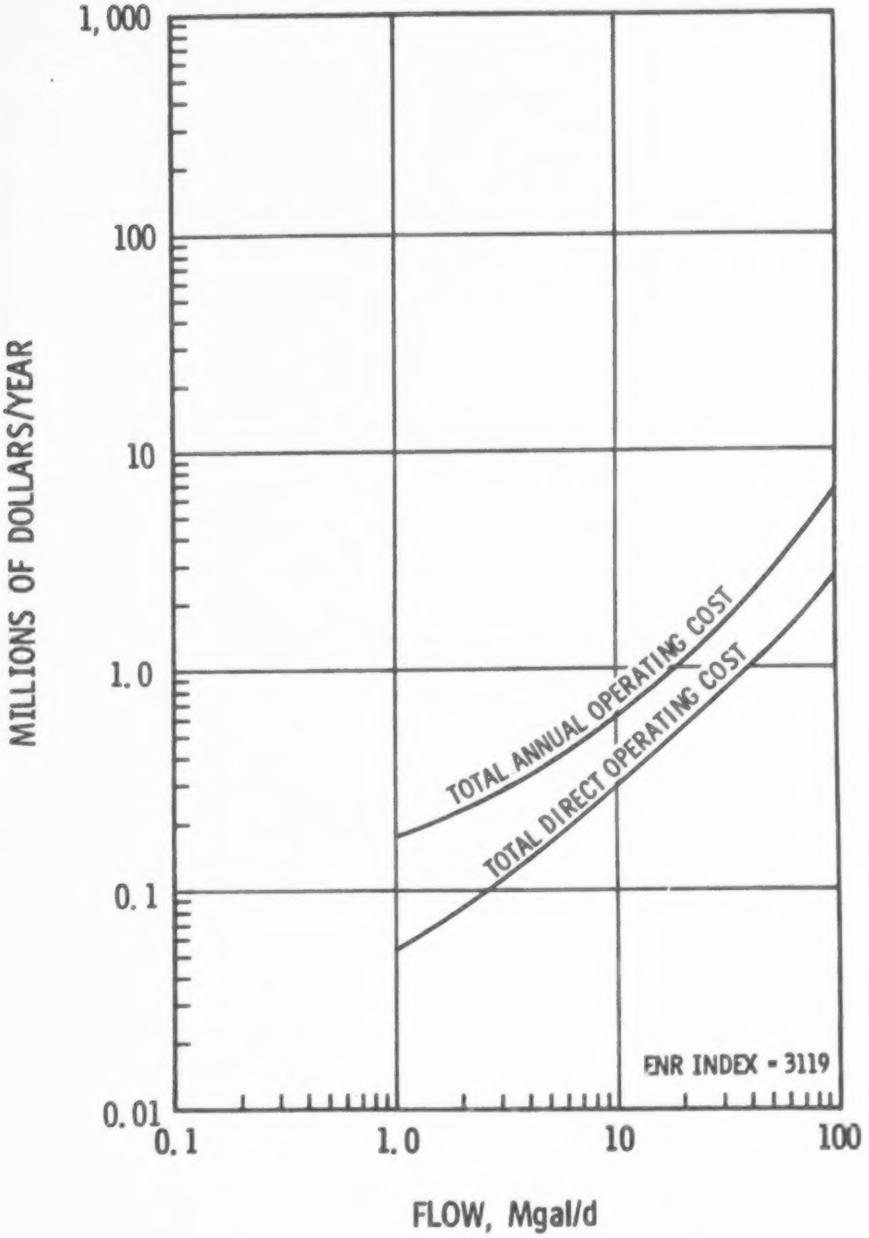
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.5.3 Lime Addition [1]

Description. Lime clarification of raw wastewater removes suspended solids as well as phosphates. There are two basic processes: the low-lime system and the high-lime system. The low-lime process consists of the addition of lime to obtain a pH of approximately 9 to 10. Generally, a subsequent biological treatment system is capable of readjusting the pH through natural recarbonation. The high-lime process consists of the addition of lime to obtain a pH of approximately 11 or more. In this case, the pH generally requires readjusting with carbon dioxide or acid to be acceptable to the secondary treatment system.

Lime can be purchased in many forms; quicklime (CaO) and hydrated lime [$Ca(OH)_2$] are the most prevalent forms. In either case, lime is usually purchased in the dry state, in bags, or in bulk. Bulk lime can be (1) shipped by trucks that are generally equipped with pneumatic unloading equipment; or (2) shipped by rail cars that consist of covered hoppers. The rail cars are emptied by opening a discharge gate, which discharges to a screw conveyor. The bulk lime is then transferred by the screw conveyor to a bucket elevator, which empties into the elevated storage tank. Bulk storage usually consists of steel or concrete bins. Storage vessels should be water- and air-tight to prevent the lime from "slaking."

Lime is generally made into a wet suspension or slurry before introduced into the treatment system. The precise steps involved in converting from the dry to the wet stage will vary according to the size of operation and type and form of limes used. In the smallest plants, bagged hydrated lime is often charged manually into a batch mixing tank with the resulting "milk-of-lime" (or slurry) being fed by means of a so-called solution feeder to the process. Where bulk hydrate is used, some type of dry feeder charges the lime continuously to either a batch or continuous mixer, then, by means of solution feeder, to the point of application. With bulk quicklime, a dry feeder is also used to feed a slaking device, where the oxides are converted to hydroxides, producing a paste or slurry. The slurry is then further diluted to milk-of-lime before being piped by gravity or pumped to the process. Dry feeders can be of the volumetric or gravimetric type.

Equipment normally associated with the lime clarification process includes bins, hoppers, conveyors and elevators, liquid storage tanks, dry and wet feeders, lime slakers, and pH instrumentation.

Typical design criteria. Typical design criteria for dissolved air flotation are discussed in Section IV.4.3.2. Typical lime requirements of a lime addition system are tabulated on the following page.

Feed water alkalinity, mg/L (as $CaCO_3$)	pH	Approximate lime dose, mg/L (as CaO)
300	9.5	185
300	10.5	270
400	9.5	230
400	10.5	380

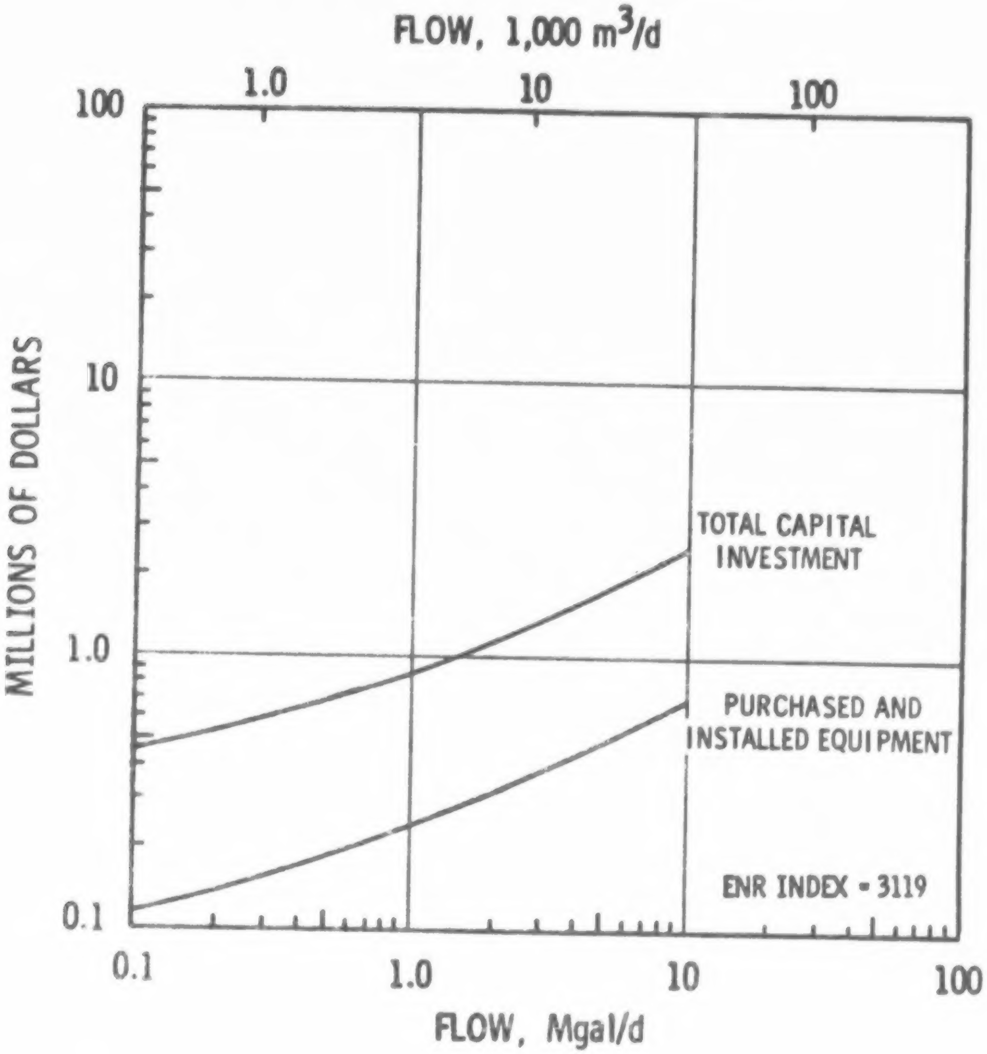
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes the DAF unit, chemical feeding equipment, chemical storage for at least 15 days at the average rate, hydrated lime for 0.1 to 10 Mgpd plant: or pebble quicklime for 10 to 100 Mgpd plants. The depreciation of equipment, piping, and buildings to house the feeding equipment are not included in the purchased and installed equipment cost. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Dosage of lime	150 mg/L
Air injection	1.25 ft^3/1,000 gal
Recycle	33%
Detention time	25 min
Air flow rate	2.3 gpm/ft^2

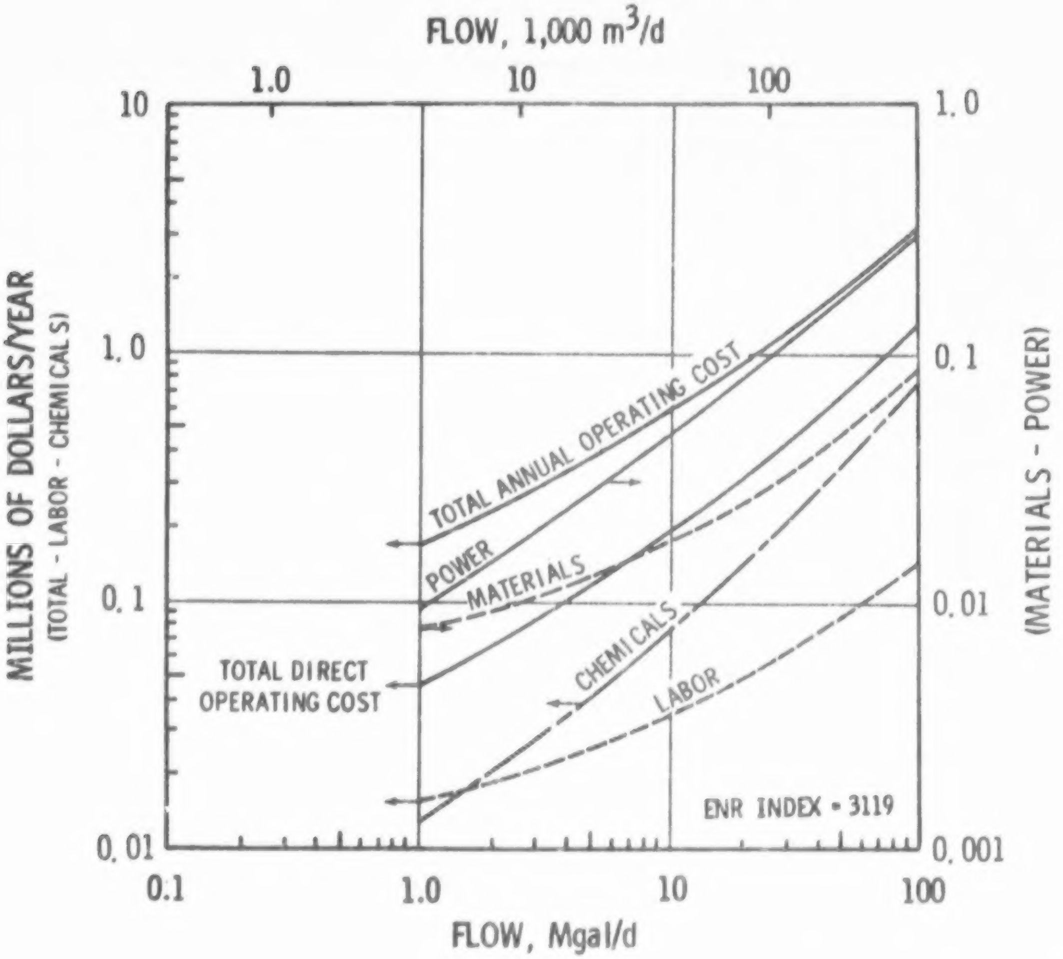
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.5.4 Polymer Addition [1]

Description. Polymers or polyelectrolytes are high-molecular-weight compounds (usually synthetic) which, when added to wastewater, can be used as coagulants, coagulant aids, filter aids, or sludge conditioners. In solution, polymers may carry either a positive, negative, or neutral charge and, as such, they are characterized as cationic, anionic, or nonionic. As a coagulant or coagulant aid, polymers act as bridges, reducing charge repulsion between colloidal and dispersed floc particles, and increasing settling velocities. As a filter aid, polymers strengthen fragile floc particles, controlling filter penetration and reducing particle breakthrough. Filterability and dewatering characteristics of sludges may similarly be improved through the use of polyelectrolytes.

Polymers are available in predissolved liquid or dry form. Dry polymers are supplied in relatively small quantities (up to about 100-lb bags or barrels) and must be dissolved on site prior to use. A stock solution, usually about 0.2 to 2.0 percent concentration, is made up for subsequent feeding to the treatment process. Preparation involves automatic or batch wetting, mixing, and aging. Stock polymer solutions may be very viscous.

Surfaces coming in contact with the polymer stock solution should be constructed of resistant materials such as type 316 stainless steel, fiberglass-reinforced plastic, or other plastic lining materials. Polymers may be supplied as a prepared stock solution ready for feeding to the treatment process. Many competing polymer formulations with differing characteristics are available, requiring somewhat differing handling procedures. Manufacturers should be consulted for optimum practices. Polymer stock solutions are generally fed to unit processes using equipment similar to that commonly in service for dissolved coagulant addition. Because of the high viscosity of stock solutions, special attention should be paid to the diameter and slopes of pipes, as well as the size of orifices used in the feed systems.

Equipment normally associated with polymer addition process includes bins, hoppers, liquid storage tanks, and dry and wet feeders.

Typical design criteria. Typical design criteria for dissolved air flotation are discussed in Section IV.4.3.2. Typical design criteria for polymer addition are shown on the next page.

Design criterion	Value
Dosage	Determined by jar testing
Materials contacting polymer solutions	Type 316 stainless steel, FRP or plastic
Storage conditions	Cool and dry; storage period should be minimized
Viscosity	Must be considered in feeding system design

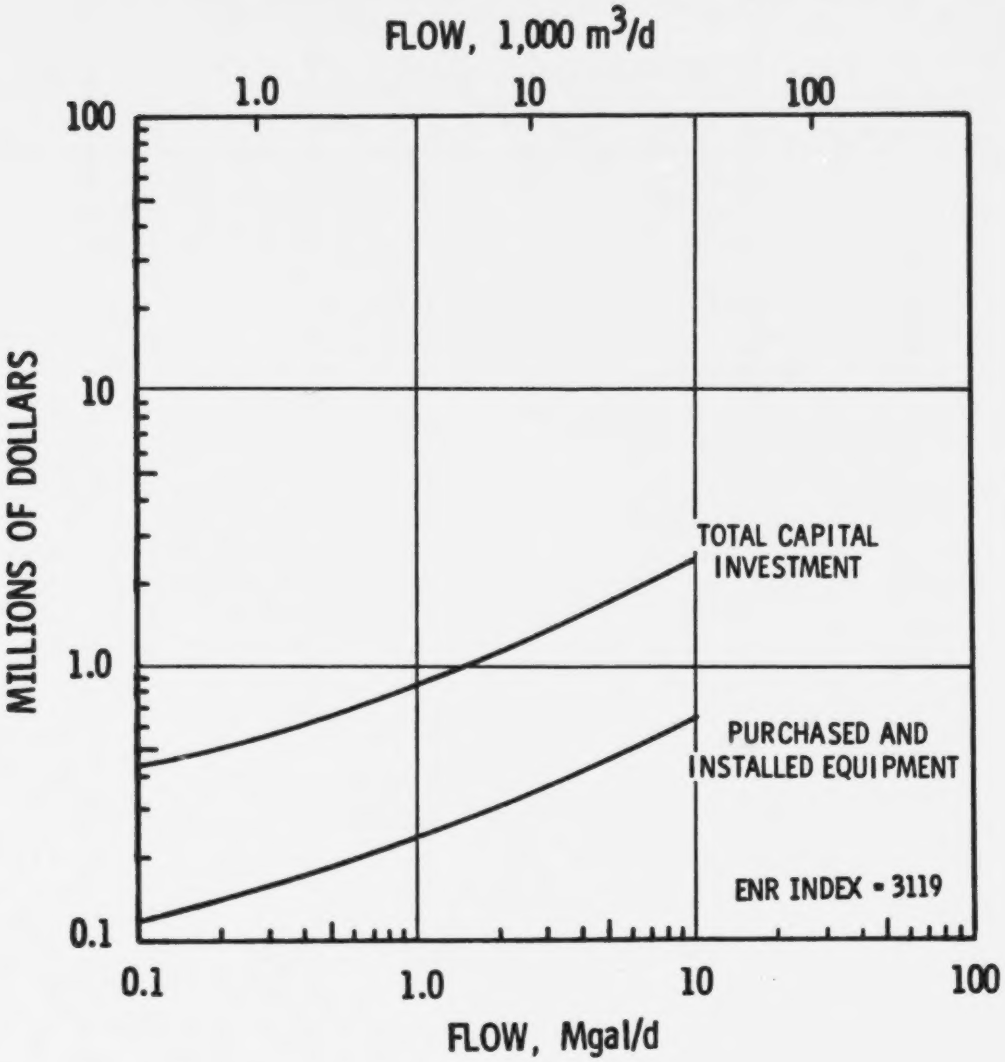
Costs. Purchased equipment and installation cost for estimating the total capital investment includes the DAF unit, chemical storage, chemical feeding, rapid mix tank, polymer solution (0.25% solution), piping and building to house the feeding equipment, and bag storage. Costs vary according to the plant size as shown below.

Plant size	Factors affecting costs
0.1 Mgpd	Cost estimation consists of manual operation feeder, mix tank, solution feeder, and holding tank; no separate building is required.
1 Mgpd and smaller	Manual procedures are applied; two systems of tanks and feeders are included in cost estimation.
10 Mgpd	Cost includes feeders and mixing tanks, two holding tanks, and 10 solution feeders; the rapid mix tank is concrete, and equipped with stainless steel mixer and handrails.

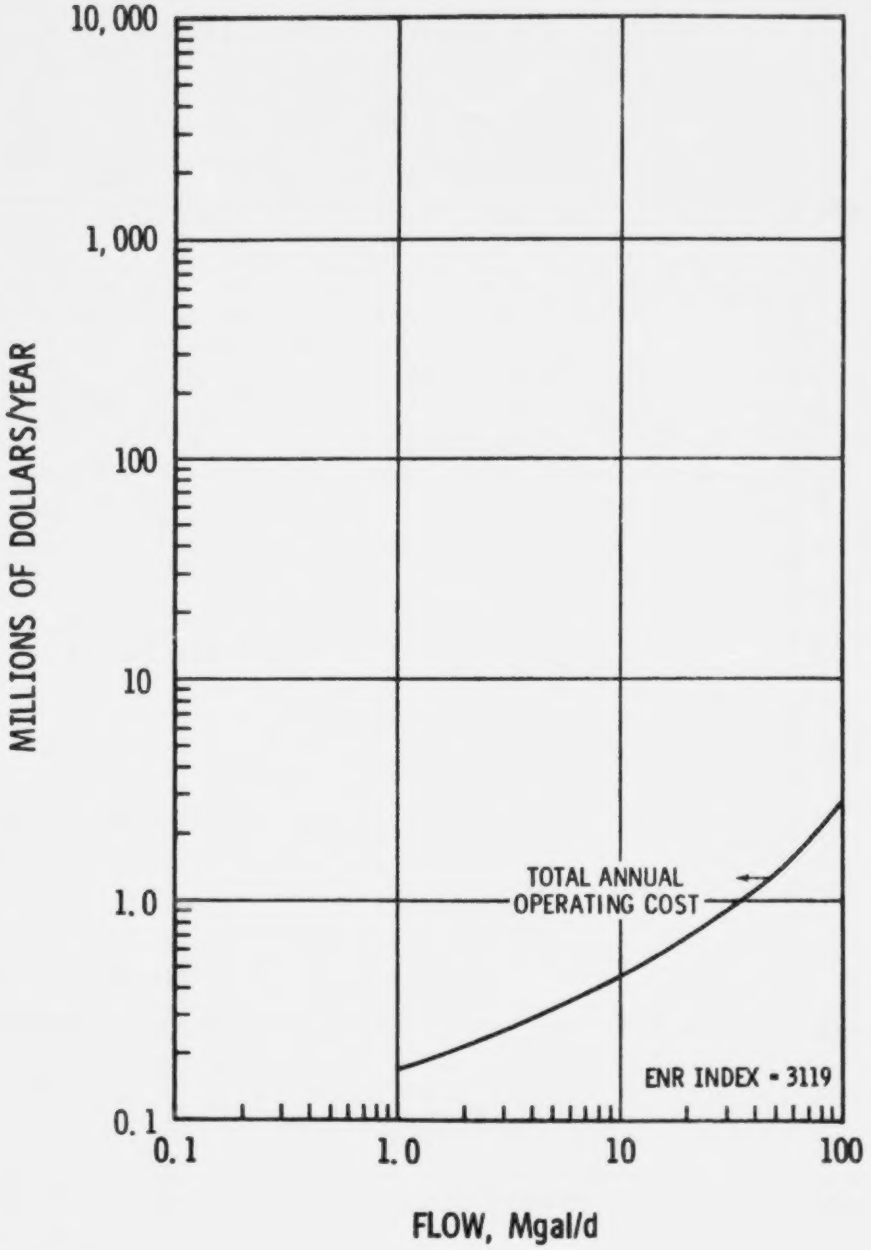
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.3.6 GRANULAR MEDIA FILTRATION [1, 2]

IV.3.6.1 Description

Granular media filtration, which is used to remove suspended solids from a liquid wastestream, is one of the oldest and most widely applied types of filtration. It utilizes a bed of granular particles (typically sand or sand with coal) as the filter medium. The bed is typically contained within a basin or tank and is supported by an underdrain system which allows the filtered liquid to be drawn off while retaining the filter medium in place. The underdrain system typically consists of metal or plastic strainers located at intervals on the bottom of the filter.

As suspended particle-laden water passes through the bed of the filter medium, particles are trapped on top of and within the bed, thus reducing its porous nature and either reducing the filtration rate at constant pressure or increasing the amount of pressure needed to force the water through the filter. If left to continue in this manner, the filter would eventually plug up with solids; the solids, therefore, must be removed. This is done by forcing a wash water stream up through the bed of granular particles. The washwater is sent through the bed at a velocity sufficiently high so that the filter bed becomes fluidized and turbulent. In this turbulent condition, the solids are dislodged from the granular particles and are discharged in the spent washwater. Air may be used initially during the backwash cycle to provide a greater surface shear force for separation of filtered particles from the granular media. This whole process is referred to as "backwashing." When the backwashing cycle is completed, the filter is returned to service.

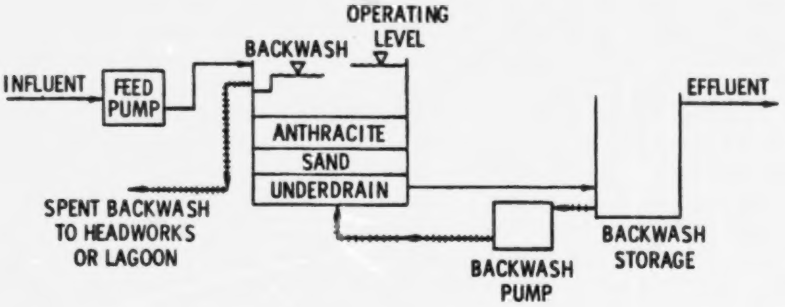
The spent backwash water contains the suspended solids removed from the liquid, and, therefore, presents a liquid disposal problem in itself. The volume of the backwash water stream, however, is normally only a small fraction (1% to 4%) of the volume of the liquid being filtered. Consequently, the suspended solids concentration of the backwash water is far greater than that of the liquid filtered. Granular media filtration essentially removes suspended solids from one liquid stream and concentrates them in another, but much smaller, liquid stream. Depending on the specific process configuration, backwash water itself can be treated to remove suspended solids by flocculation and/or sedimentation or by returning it to the portion of the process from whence the liquid stream subjected to filtration originated.

Dual-media filtration involves the use of both sand and anthracite as filter media, with anthracite being placed on top of the sand. This provides a large pore size near the filter surface allowing a large degree of solids capture with a relatively low head loss. The smaller pores in the lower sand media act as a polishing unit to remove the smaller sized particles from the wastewater. For wastewater filtration involving relatively high suspended solids concentrations compared to potable water treatment, a gradation in media pore size from large to small in the direction of flow is required to provide a feasible length of filter run prior to backwashing. Dual-media, multimedia and upflow immedium filters serve this purpose.

Filtration systems can be constructed of concrete or steel, with single or multiple compartment units. Steel units can be either horizontal or vertical and are generally used for pressure filters. Systems can be manually or automatically operated.

Backwash sequences can include air scour or surface wash steps. Backwash water can be stored separately or in chambers that are integral parts of the filter unit. Backwash water can be pumped through the unit or can be supplied through gravity head tanks.

Equipment normally associated with the dual-media filtration process includes dual-media filters, blowers, and controls. A flow diagram of a dual-media filtration system is shown below.



IV.3.6.2 Typical Design Criteria

Typical design criteria for a dual-media filtration system are tabulated below.

Design criteria	Value
Filtration rate	2 to 8 gpm/ft^2
Bed depth	24 to 48 in.
Depth ratio	1.1: to 1:4 (sand to anthracite)
Backwash rate	15 to 25 gpm/ft^2
Air scour rate	3 to 5 standard ft^3/min/ft^2
Filter run length	8 to 48 hr
Terminal head loss	6 to 15 ft

Note: Multi-media filtration utilizes similar criteria; however, the depth ratio will differ.

IV.3.6.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes facilities for backwash storage, all feed and backwash pumps, piping, and building. The following operating characteristics were assumed for cost estimation of a dual-media filtration system:

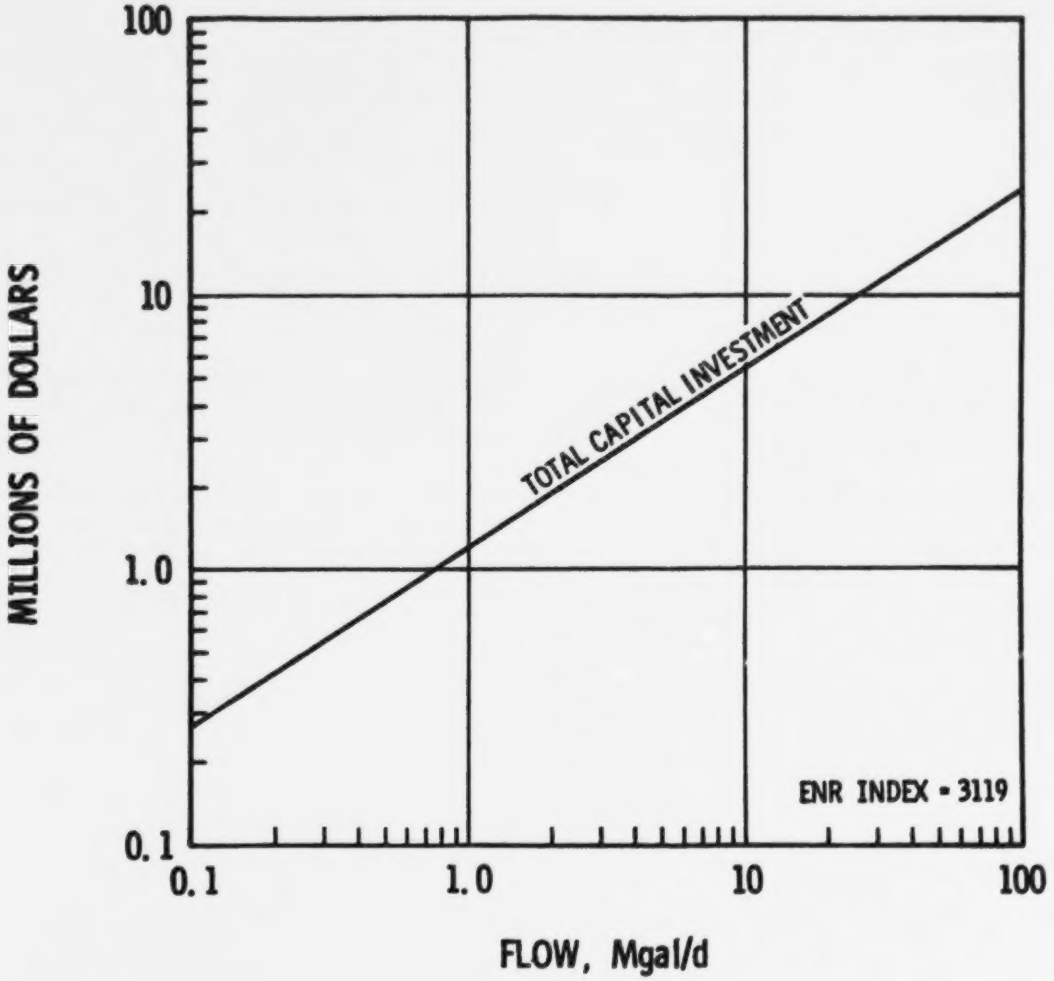
Operating characteristic	Assumed value
Gravity filters	4 gpm/ft^2
TDH for backwash and feed pumps	14 ft
Run length	12 hr; 15 min backwash at 15 gpm/ft^2
Pump efficiency	70%
Motor efficiency	93%
Centrifugal pumps	-
Backwash holding tank	Capacity of two backwash cycles

IV.3.6.4 References

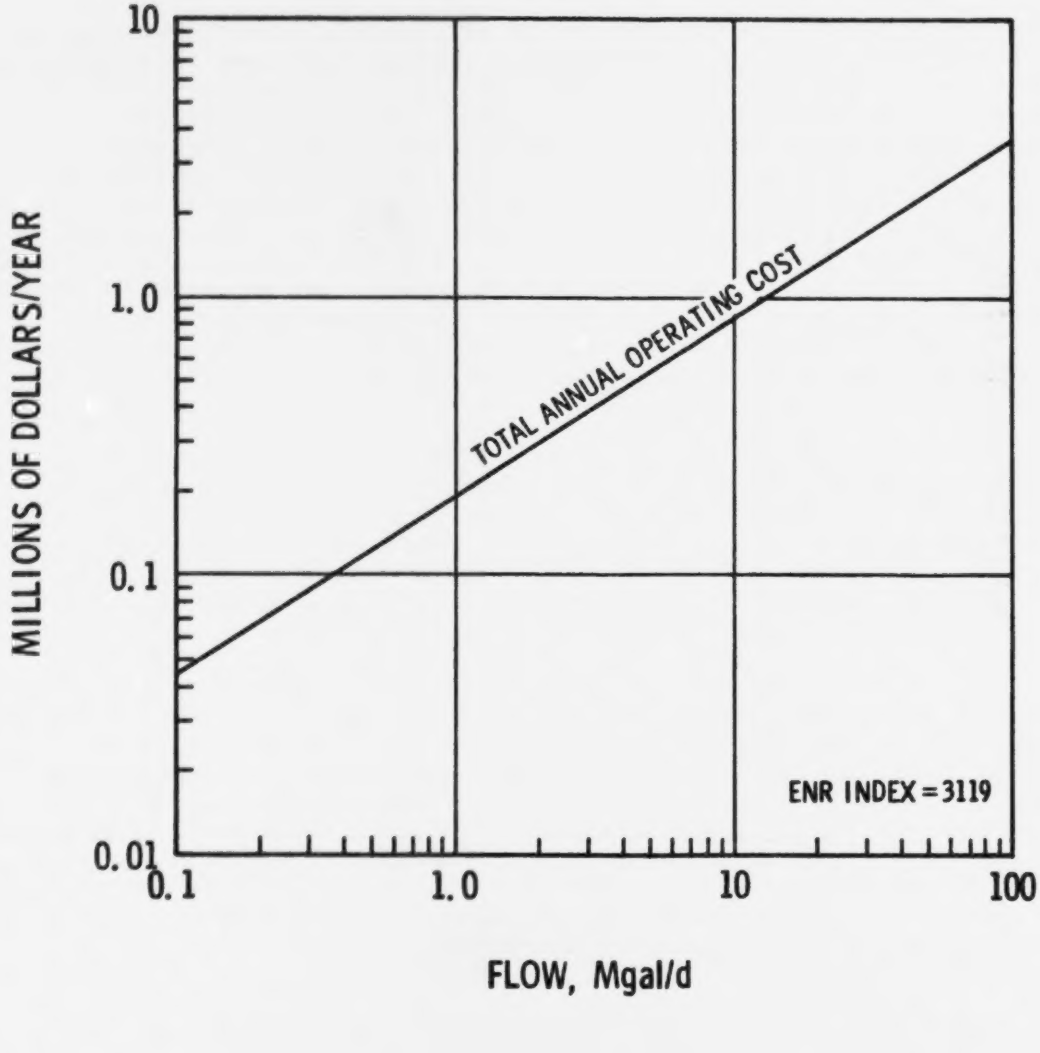
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

2. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 22-1-22-25.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



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IV.3.7 ULTRAFILTRATION [1,2]

IV.3.7.1 Description

Ultrafiltration is a membrane filtration process that separates high-molecular-weight solutes or colloids from a solution or suspension. The process has been successfully applied for both homogeneous solutions and colloidal suspensions, which are difficult to separate practically by other techniques. To date, commercial applications have been entirely focused on aqueous media.

The basic principle of operation of ultrafiltration can be explained as follows: Flowing by a porous membrane is a solution containing two solutes: one solute has a molecular size too small to be retained by the membrane, and the other has a larger size that allows 100% retention. A hydrostatic pressure, typically 10 to 100 psig, is applied to the upstream side of the supported membrane, and the large-molecule solute or colloid is retained (rejected) by the membrane. A fluid concentrated in the retained solute is collected as a product from the upstream side; a solution of small-molecule solute and solvent is collected from the downstream side of the membrane. Of course, where only a single solute is present and is rejected by the membrane, the liquid collected downstream is (ideally) pure solvent.

Retained solute (or particle) size is one characteristic distinguishing ultrafiltration from other filtration processes. Viewed on a spectrum of membrane separation processes, ultrafiltration is only one of a series of membrane methods that can be used. For example, reverse osmosis, a membrane process capable of separating dissolved ionic species from water, falls further down the same scale of separated particle size.

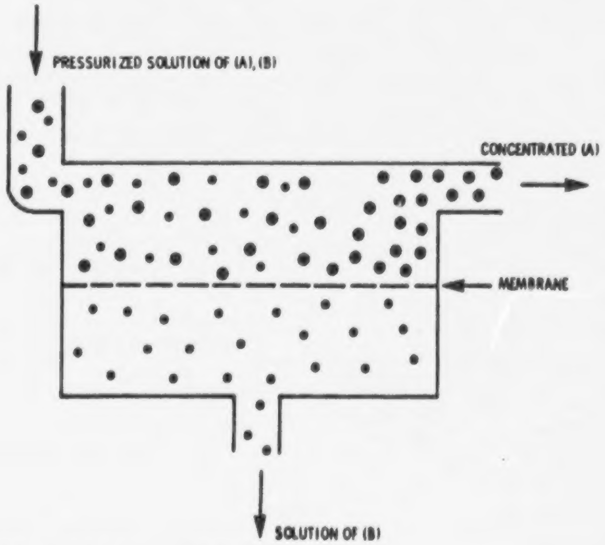
Ultrafiltration membranes are asymmetric structures, possessing an extremely thin selective layer (0.1- to 1.0-μm thick) supported on a thicker spongy substructure. Controlled variation of fabrication methods can produce membranes with desirable retentitive characteristics for a number of separation applications. It has become possible to tailor membranes with a wide range of selective properties. For example, tight membranes can retain organic solutes of 500 to 1,000 molecular weight while allowing passage of most inorganic salts; conversely, loose membranes can discriminate between solutes of 1,000,000 versus 250,000 molecular weight.

Ultrafiltration membranes are different from so-called "solution-diffusion" membranes, which have been studied for a wide variety of gas and liquid-phase separations. The latter group possesses a permselective structure that is nonporous, and separation is effected on the basis of differences in solubility and molecular diffusivity within the actual polymer matrix. Reverse osmosis membranes generally fall into this category.

Membranes can be made from various synthetic or natural polymeric materials. These range from hydrophilic polymers, such as cellulose, to very hydrophobic materials, such as fluorinated polymers. Polyarylsulfones and inorganic materials have been introduced to deal with high temperatures and pH values.

Membranes of this type are in many respects similar to reverse osmosis membranes except for the openness of their pores. Other forms and materials are available as well, including porous zirconia deposited on a porous carbon substrate and on a porous ceramic tube. The latter two systems, while more expensive than the former, are capable of use to very high pH values and temperatures.

A flow diagram of an ultrafiltration process is shown below.



IV.3.7.2 Typical Design Criteria

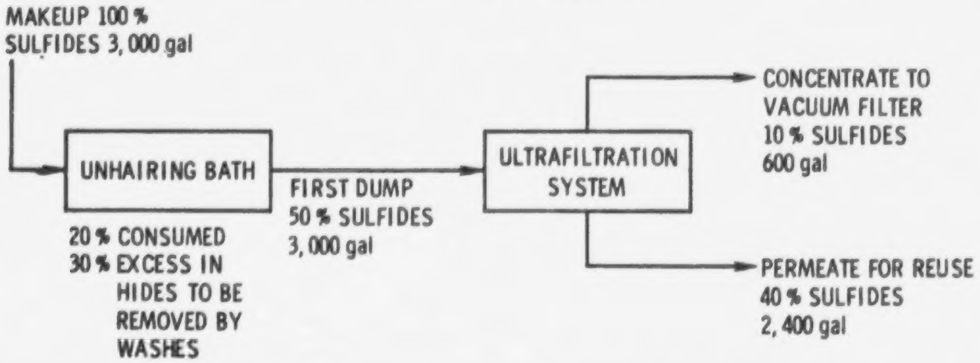
Design criteria	Value
Membrane:	
Material	Most organic polymers
Water permeability	7-290 gpd per sq ft at 30 psi
Molecular weight	340-45,000
Retentivity	60-100 percent
Maximum operating temperature	50-120°C
Driving force	Approximately 25 psi

IV.3.7.3 Costs

Capital and operating costs for ultrafiltration are strongly dependent on the specific application and the capacity of the system; there is no typical model. For the purpose of cost estimation, data are generated for sulfide recovery by the ultrafiltration method. Design basis for economic projection of sulfide recovery by ultrafiltration are shown on the next page.

Category	Item	Basis
Tannery operation	Tannery output	50,000 lb of hide/d
	Discharge treated	First dump only
	Volume treated	3,000 gpd (25,000 lb water)
	Hair burn liquor	3% sulfhydrate 1% sulfide flakes 4% lime (percentages based on hide weight)
	Operating period	260 d/yr
Ultrafiltration system	Membrane flux	40 gal/ft^2-d
	Water recovery	80% (2,400 gpd)
	Membrane life	3 yr

A flow diagram of sulfide recovery by ultrafiltration, including operating characteristics, is shown below.


MAKEUP 100 %
SULFIDES 3,000 gal
UNHAIRING BATH
20 % CONSUMED
30 % EXCESS IN HIDES TO BE REMOVED BY WASHES
FIRST DUMP
50 % SULFIDES
3,000 gal
ULTRAFILTRATION SYSTEM
CONCENTRATE TO VACUUM FILTER
10 % SULFIDES
600 gal
PERMEATE FOR REUSE
40 % SULFIDES
2,400 gal

Fixed capital cost is approximately $136,000 for a daily feed volume of 0.003 Mgpd. Total capital cost is estimated to be $144,000. Estimation of annual operating cost is presented in the following table, found on the next page.

Cost item	Annual quantity	Cost per unit quantity	Annual cost,[a] $	
Direct operating cost				
Labor				
Operating	520 man-hr	$10/hr	5,200	
Maintenance	52 man-hr	$12/hr	620	
				5,820
Chemicals (cleaning)[b]				430
Membrane replacement			900	
Maintenance			1,250	
				2,150
Power	18,200 kWh	$0.035/kWh		640
Total direct operating cost				9,040
Total indirect operating cost				23,100
Total annual operating cost				32,100

[a] Design basis included 260 d/yr, 3,000 gpd, 80% water recovery, 60-ft^2 membrane area, and $40,000 purchased equipment cost.

[b] 200 gal/wk x 8.33 lb/gal x 0.5% chemicals x $1/lb÷5 d/wk x 260 d/yr.

Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying flowrates:

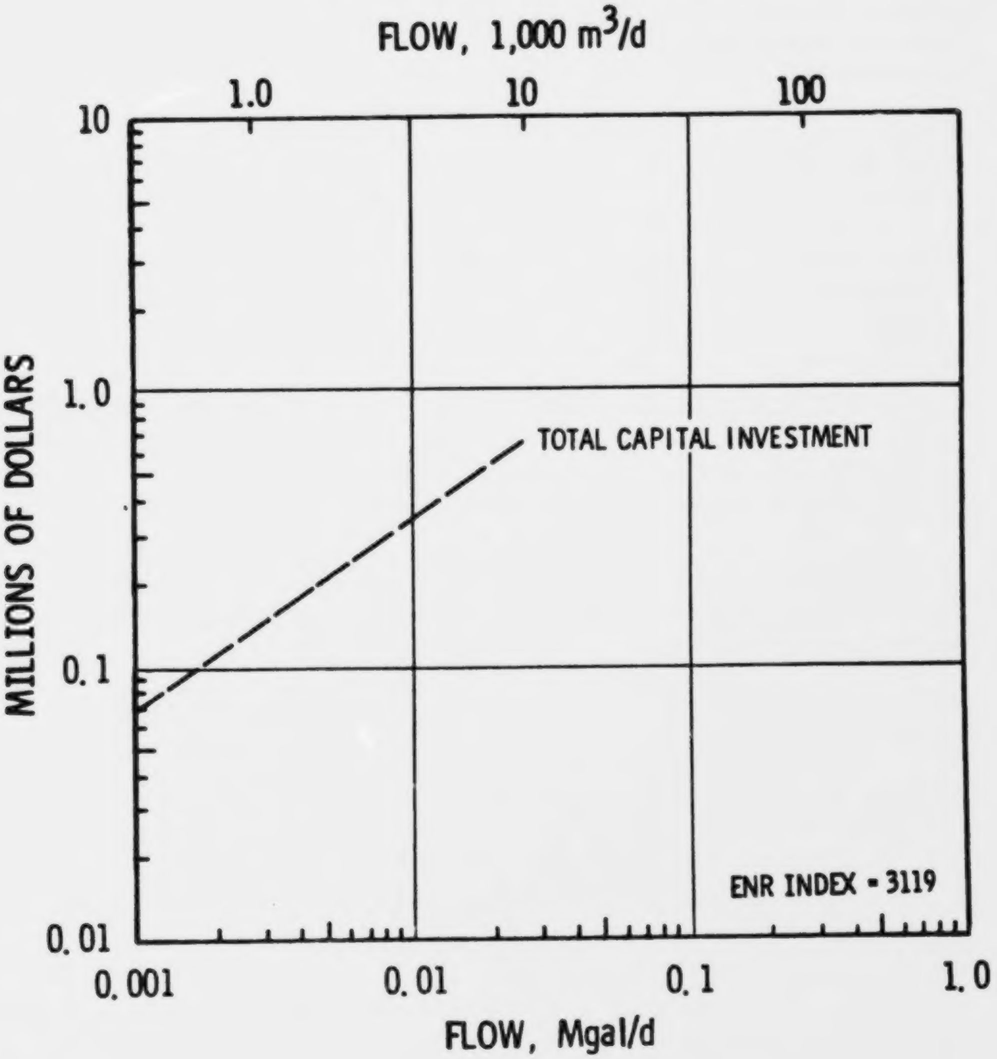
Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Chemicals, power and materials	1.0

IV.3.7.4 References

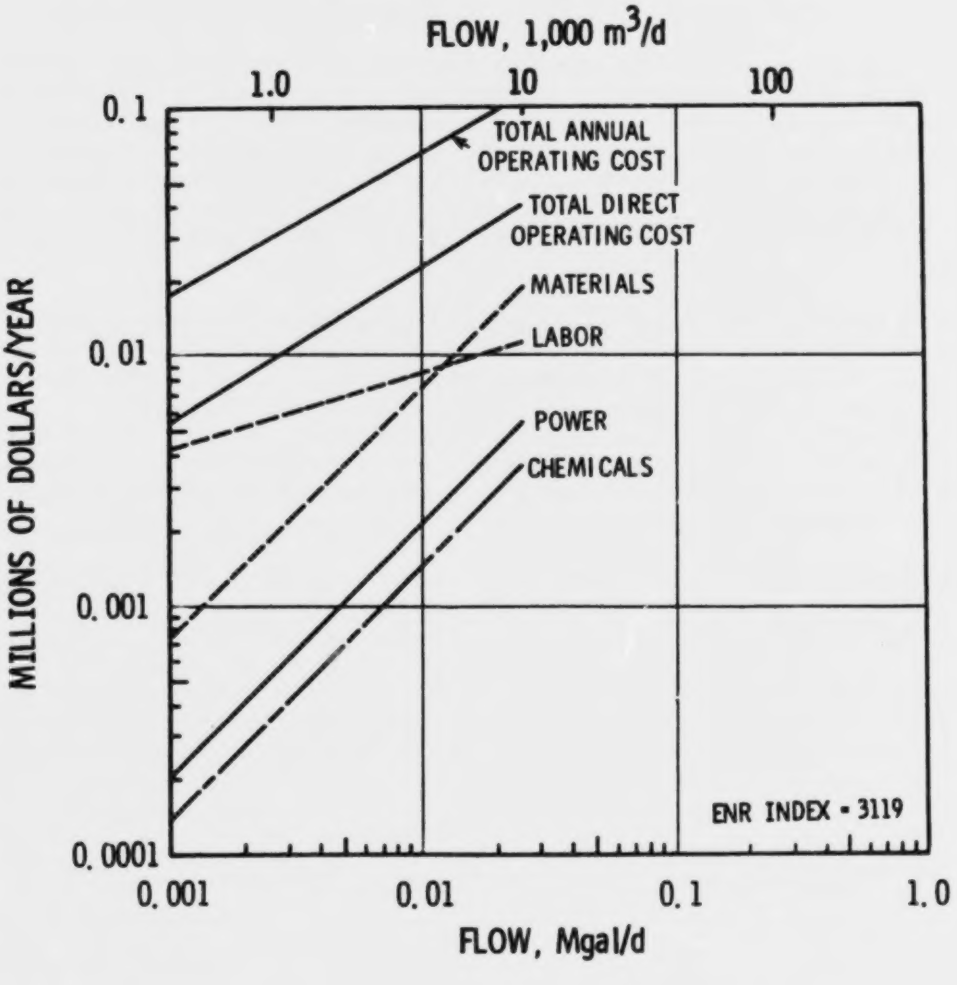
1. Abcor, Inc. A New Approach for Treatment of Spent Tannery Liquors, Massachusetts 1979.

2. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 33-1 to 33-13.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.4 SECONDARY WASTEWATER TREATMENT

IV.4.1 ACTIVATED SLUDGE

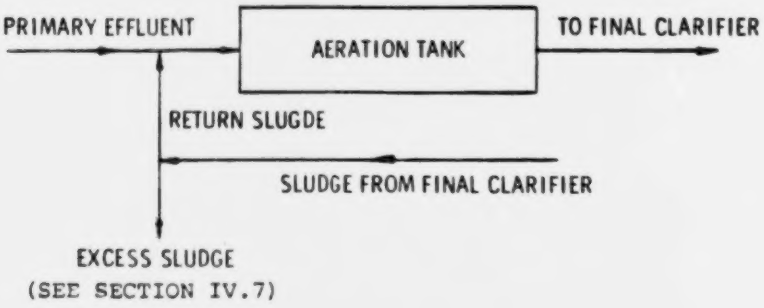
IV.4.1.1 Conventional Activated Sludge (Diffused Aeration) [1]

Description. In the conventional activated sludge plant, the wastewater is commonly aerated in a plug-flow hydraulic mode. Diffusers are employed to transfer oxygen from air to wastewater. Compressors are used to supply air to the submerged systems, normally through a network of diffusers, although newer submerged devices which do not come under the general category of diffusers (e.g., static aerators and jet aerators) are being developed and applied.

Diffused air systems may be classified as fine bubble or coarse bubble. Diffusers commonly used in activated sludge service include the following: porous ceramic plates laid in the basin bottom (fine bubble), porous ceramic domes or ceramic or plastic tubes connected to a pipe header and lateral system (fine bubble), tubes covered with synthetic fabric or wound filaments (fine or coarse bubble), and specially designed spargers with multiple openings (coarse bubble).

Common modifications include step aeration, contact stabilization, and complete mix flow regimes. Aluminum or ferric chloride is sometimes added to the tank for phosphorus removal.

Equipment normally associated with diffused air, activated sludge systems includes air diffusers and compressors. A flow diagram for conventional activated sludge with diffused aeration is shown below.



Typical design criteria. A partial listing of typical design criteria for the conventional activated sludge process is summarized below.

Design criteria	Value
Mixed liquor suspended solids (MLSS)	1,500-3,000 mg/L
Food-to-microorganism (F/M) ratio	0.25-0.5 lb BOD_5/d/lb MLSS
Air required	800-1,500 standard ft^3/lb BOD_5 removed
Sludge retention time	5-10 d

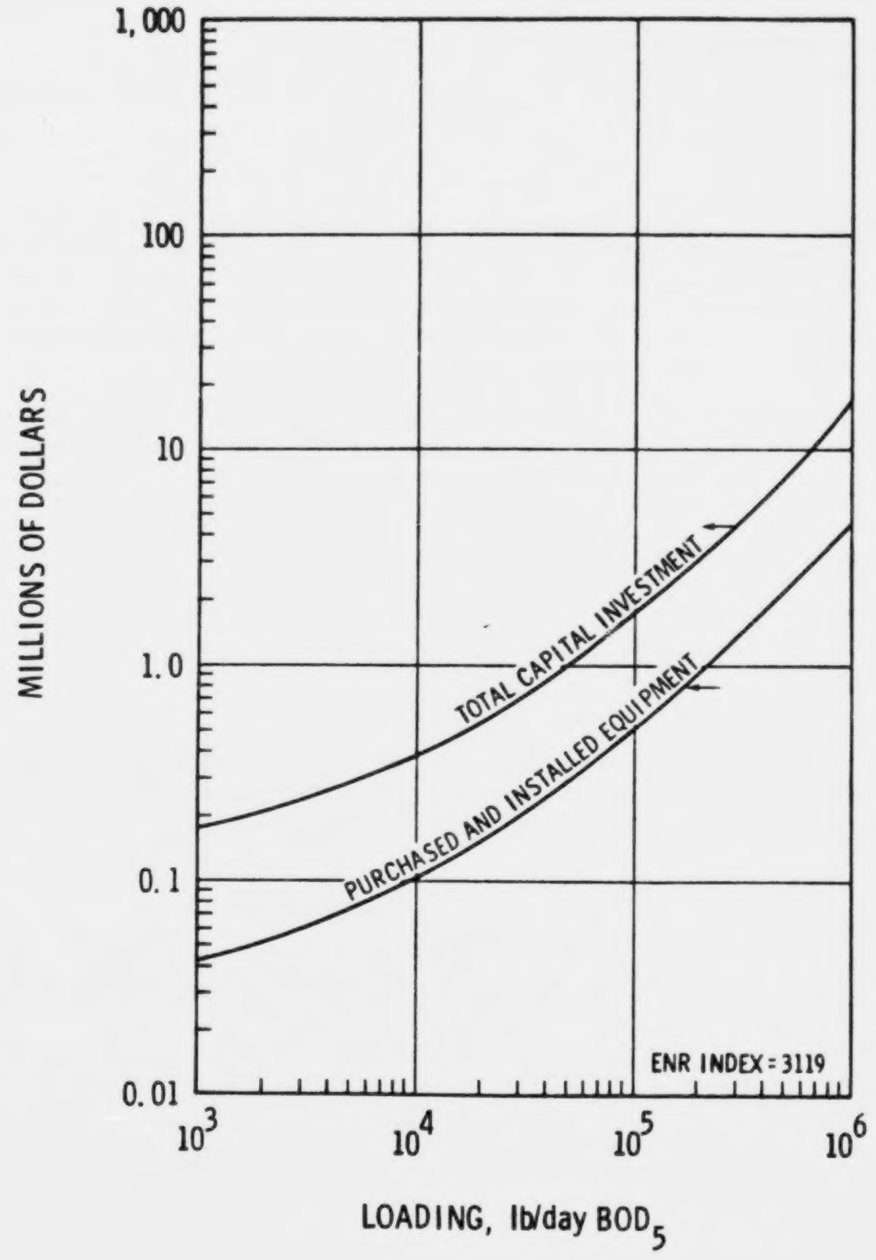
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes aeration basins, air supply and dissolution equipment and piping, and blower building. Clarifier and recycle pumps are not included in the purchased and installed equipment cost. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Organic loading	1,300 mg/L BOD_5
Service life	40 yr
Oxygen requirement	1.1 lb O_2 supplied/lb BOD_5 removed
MLSS	2,000 mg/L
F/M ratio	0.25 lb/BOD_5/d/lb MLSS

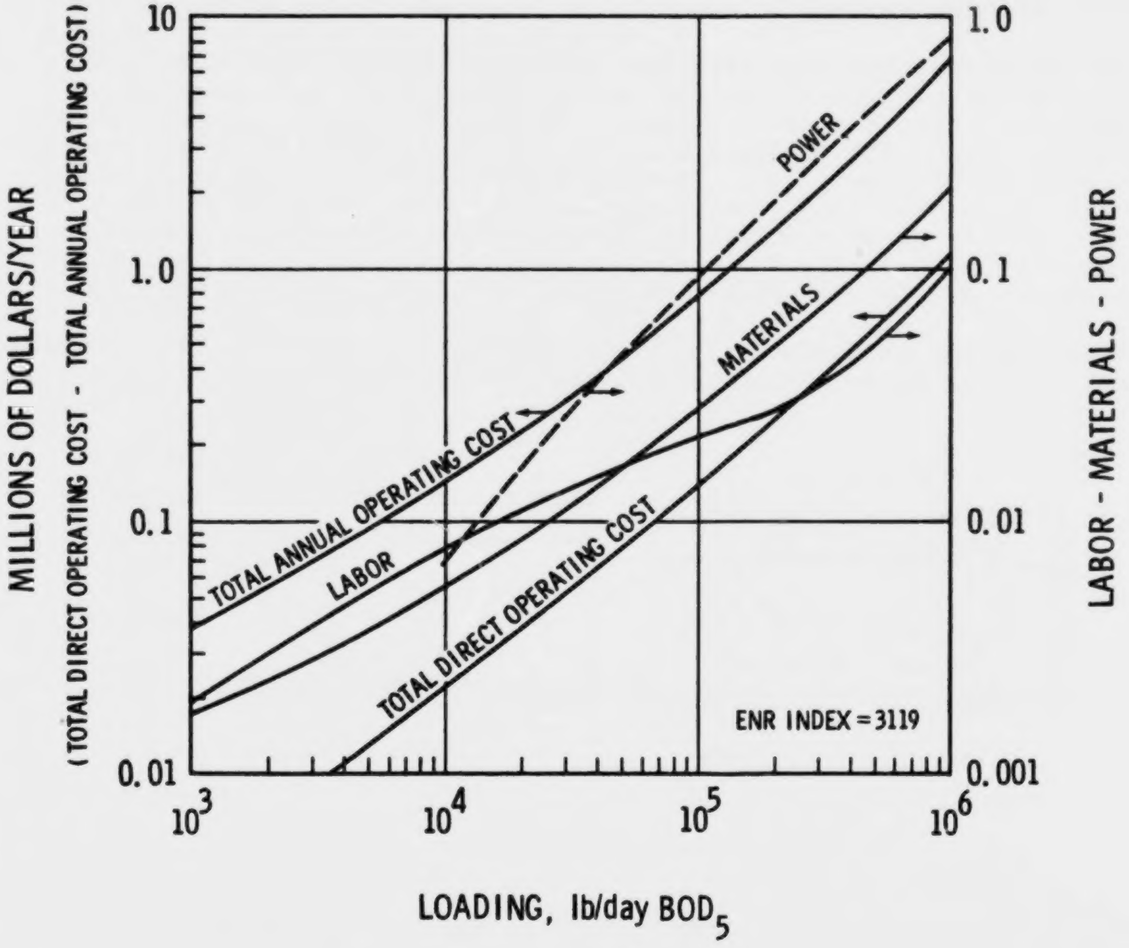
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST

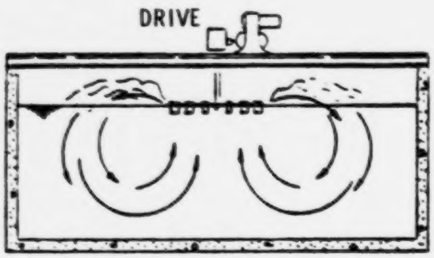


IV.4.1.2 Conventional Activated Sludge (Mechanical Aeration) [1]

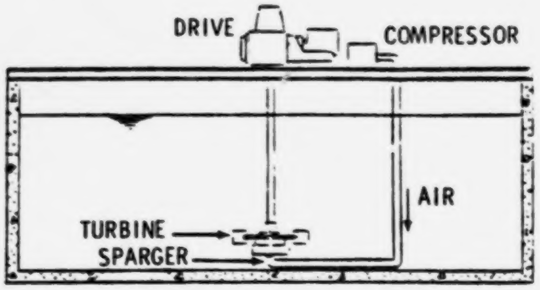
Description. Mechanical aeration methods include the surface-type mechanical entrainment aerators and the submerged turbine with compressed air spargers (agitator/sparger system). The surface-type aerators entrain atmospheric air by producing a region of intense turbulence at the surface around their periphery. They are designed to pump large quantities of liquid, thus dispersing the entrained air and agitating and mixing the basin contents. The agitator/sparger system consists of a radial-flow turbine located below the mid-depth of the basin with compressed air supplied to the turbine through a sparger. Volatile compounds are removed to a certain extent in the aeration process. Metals will also be partially removed, with accumulation in the sludge.

The submerged turbine aeration system affords a convenient and relatively economical method for sludge plants that have high capacity input per unit volume. To attain optimum flexibility of oxygen input, the surface aerator can be combined with the submerged turbine aerator. Several manufacturers supply such equipment, with both aerators mounted on the same vertical shaft. Such an arrangement might be advantageous if space limitations require the use of deep aeration basins. In addition, mechanical aerators may be either the floating or fixed installation type.

Diagrams for activated sludge sytems containing mechanical surface aerators or submerged turbine aerators are shown below.



MECHANICAL SURFACE AERATOR



SUBMERGED TURBINE AERATOR

Typical design criteria. A partial listing of typical design criteria for the conventional activated sludge process is summarized below.

Design criteria	Value
Mixed liquor suspended solids (MLSS)	1,500-3,000 mg/L
Food-to-microorganism (F/M) ratio	0.25-0.5 lb BOD_5/d/lb MLSS
Sludge age	5-10 d

The mixing equipment for aeration or oxygen transfer must be sized to keep the solids in uniform suspension at all times. Depending on basin shape and depth, 4,000 mg/L of MLSS require about 0.75 to 1.0 hp/1,000 ft^3 (0.02 to 0.03 kW/m^3) of basin volume to prevent settling if mechanical aerators are employed. However, the power required to transfer the necessary oxygen will usually equal or exceed this value.

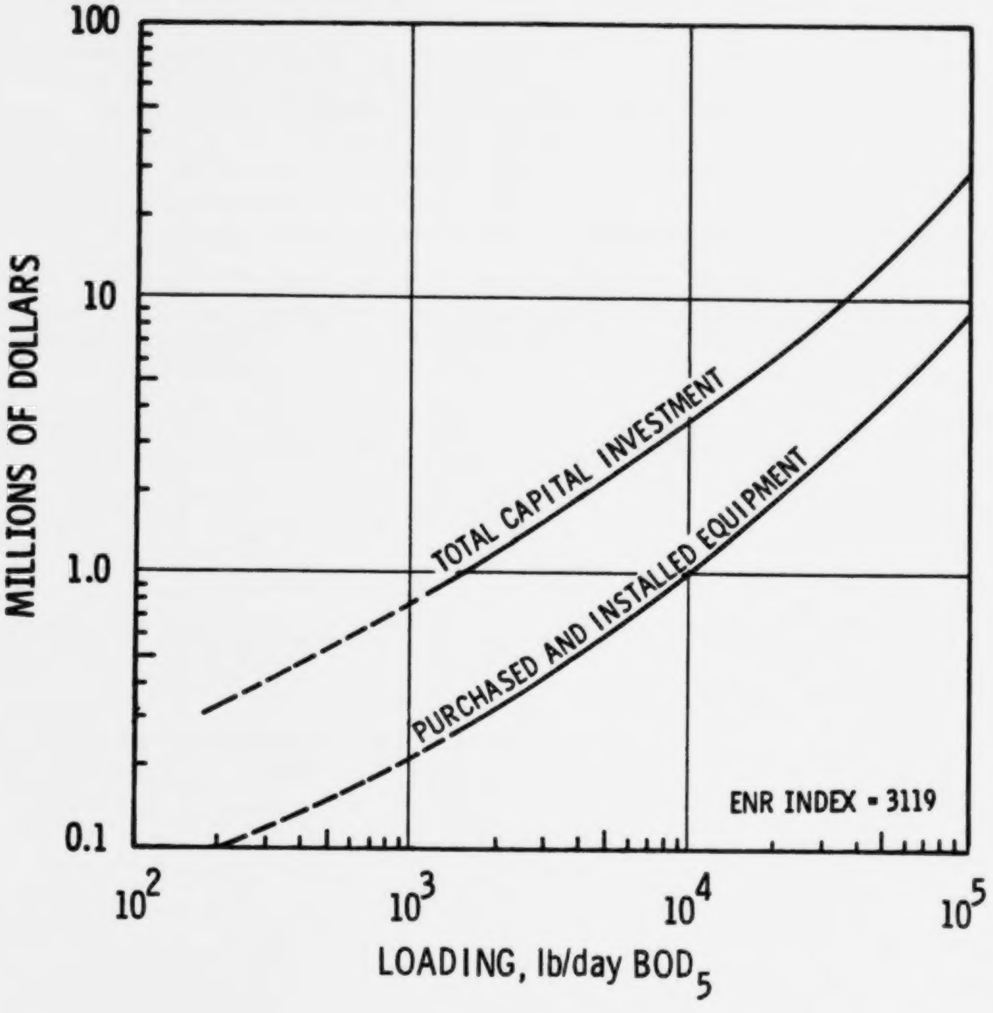
Costs. Purchased equipment and installation cost for estimation of the total capital investment includes aeration basins. Clarifier and recycle pumps are not included. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Organic loading	1,200 mg/L BOD_5
Oxygen requirement	1.1 lb O_2 supplies/lb BOD_5 removed
MLSS	2,000 mg/L
F/M	0.25 lb BOD_5/d/lb MLSS

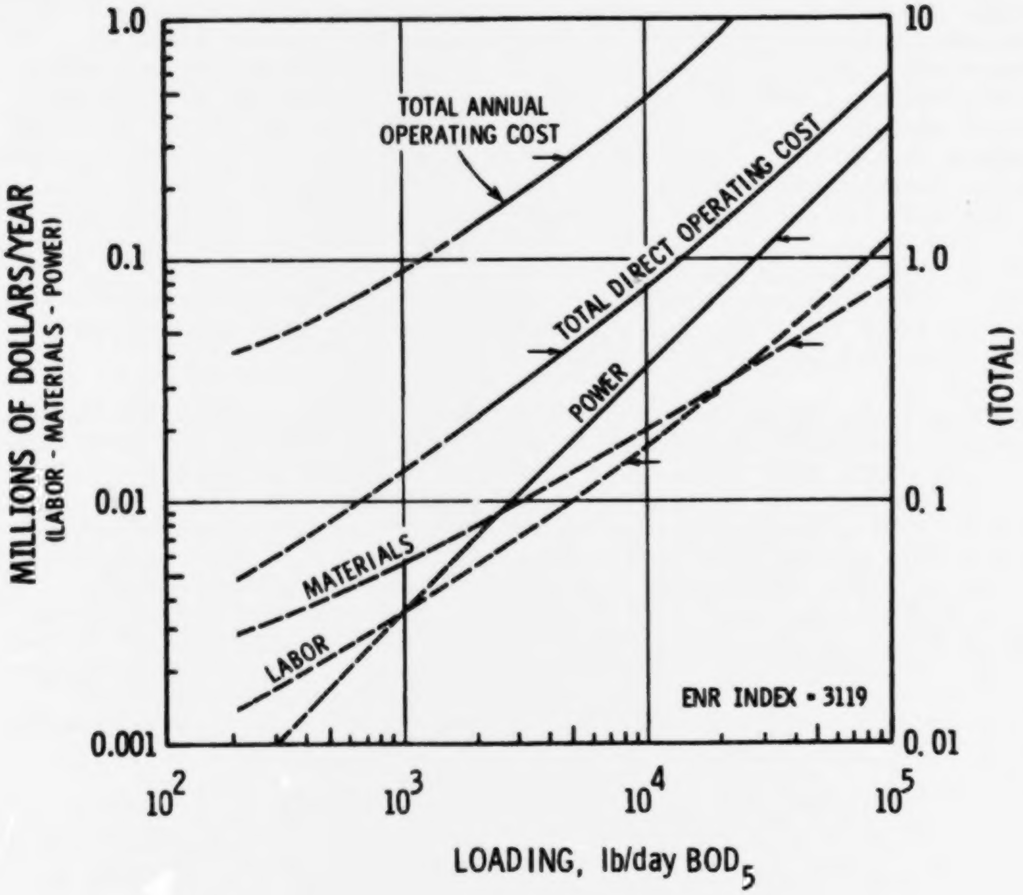
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



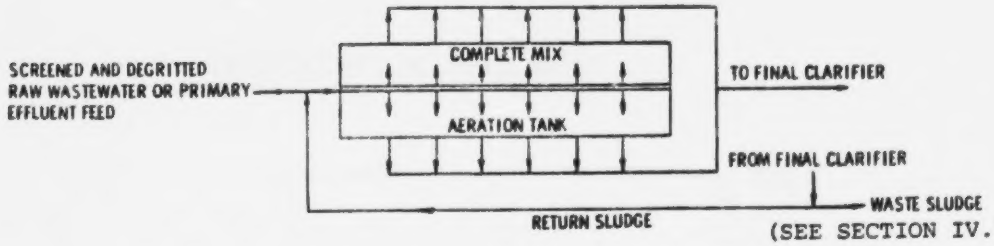
IV.4.1.3 High-rate Activated Sludge-Diffused Aeration [1]

Description. A description of the activated sludge process in general is presented in Section IV.4.1.1. Activated sludge systems have traditionally been classified as high-rate, conventional, or extended aeration based on organic loading.

High-rate air-activated sludge systems are designed with a food-to microorganisms (F/M) loadings in the range of 0.75 to 1.5 lb BOD_5/d/lb MLVSS (mixed liquor volatile suspended solids). Moreover, they are usually characterized by mixed liquor suspended solids - not necessarily (MLSS) concentrations, short aeration detention times, high volumetric loadings, low air usage rates, and intermediate levels of BOD_5 and suspended solids removal efficiencies. Prior to enactment of nationwide secondary treatment regulations, this technology was utilized as an independent treatment system for plants where BOD_5 removals of 50 to 70 percent would suffice. With present-day treatment requirements, it no longer qualifies as a "stand-alone" activated sludge option. These systems are normally designed to operate in either complete-mix or plug-flow hydraulic configurations. Either surface or submerged aeration systems can be employed to transfer oxygen from air to wastewater, although submerged equipment is specified more frequently for this process. Compressors are used to supply air to submerged aeration systems.

Due primarily to rapidly escalating power costs, interest has been recently expressed in the development of high-rate, diffused aeration systems that would produce a high quality secondary effluent. The key to development of efficient high-rate air systems is the availability of submerged aeration equipment that could satisfy the high oxygen demand rates that accompany high MLSS levels. New innovations in fine bubble diffuser and jet aeration technology offer potential for uniting high-efficiency oxygen transfer with high-rate, air-activated sludge-flow regimes to achieve acceptable secondary treatment as independent "stand-alone" processes. Research evaluations and field studies currently underway should provide performance and cost data on this subject in the next several years.

Equipment normally associated with high-rate activated sludge systems include air diffusers and compressors. A flow diagram of the high-rate activated sludge system is shown on the next page.



(SEE SECTION IV.7)

Typical design criteria. A partial listing of typical design criteria for the modified aeration process and the high-solids, high-rate aeration process is summarized below.

Design criteria	Units	Modification	High solids, high-rate aeration[a]
Mixed liquor suspended solids (MLSS)	mg/L	800 - 2,000	3,000 - 5,000
Food-to-microorganism (F/M) ratio	lb BOD_5/d/lb MLVSS[b]	0.75 - 1.5	0.4 - 0.8
Air required	standard ft^3 air/lb BOD_5 removed	400 - 800	800 - 1,200
Oxygen required	lb O_2/lb BOD_5 removed	0.4 - 0.7	0.9 - 1.2
Sludge age	days	0.75 - 2	2 - 5
Recycle ratio		0.25 - 1.0	1.0 - 5.0
Volatile fraction of MLSS		0.7 - 0.85	0.7 - 0.8

[a] Tentative.

[b] MLVSS = mixed liquor volatile suspended solids.

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes aeration basins, air supply equipment and piping, and a blower building. Clarifier and recycle pumps are not included in the purchased and installed equipment cost. Basins are sized with 50% recycle flow. The following operating characteristics were assumed for cost estimation:

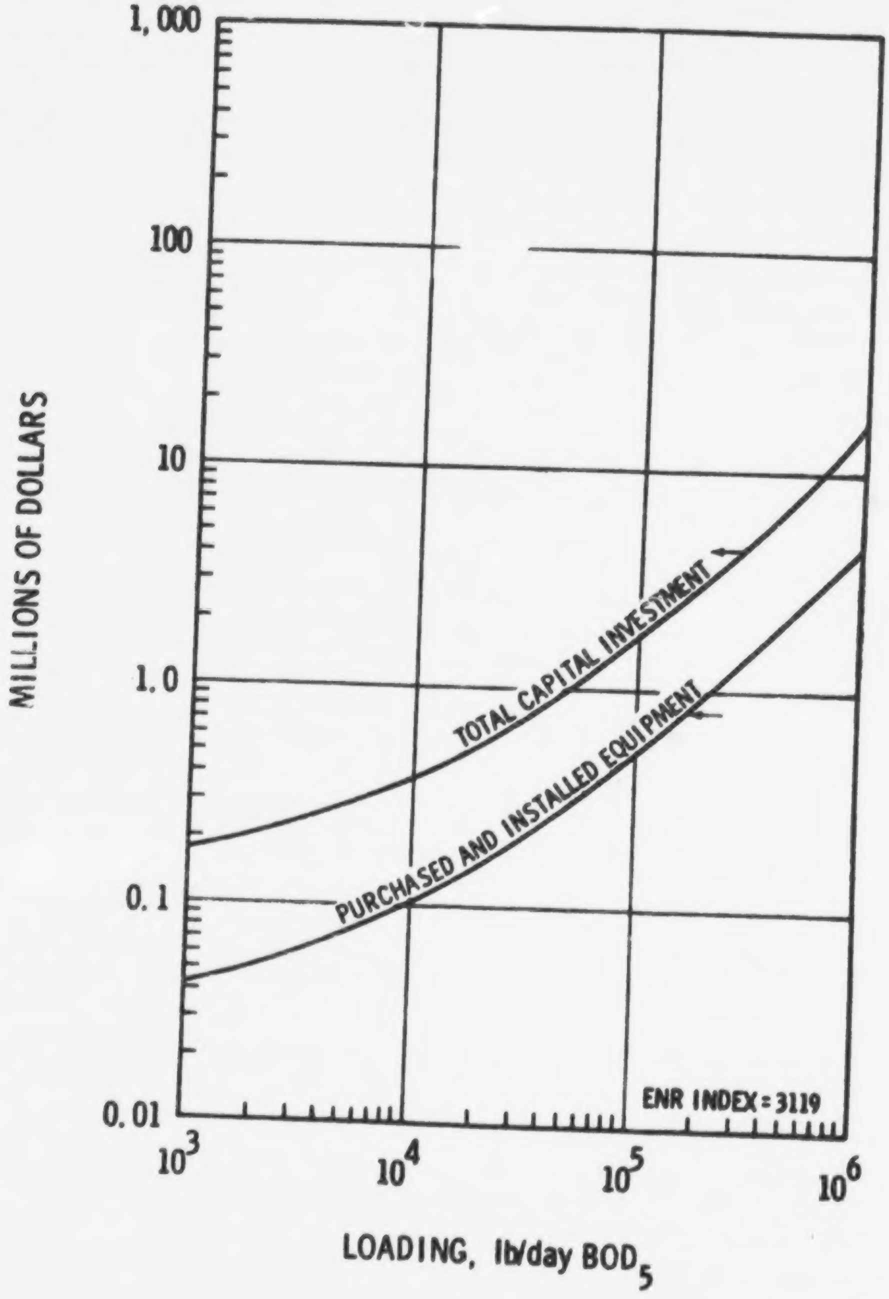
Operating characteristics	Assumed value
Organic loading	1,300 mg/L BOD_5
F/M	1.0 lb BOD_5/d/lb MLVSS
Oxygen requirement	0.7 lb BOD_5 removed
MLVSS	1,050 mg/L
Service life	40 yr

References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

2. Metcalf & Eddy, Inc., Wastewater Engineering, Collection Treatment Disposal, McGraw-Hill Publishing Co., New York, New York, 1972. 782 pp.

TOTAL CAPITAL INVESTMENT



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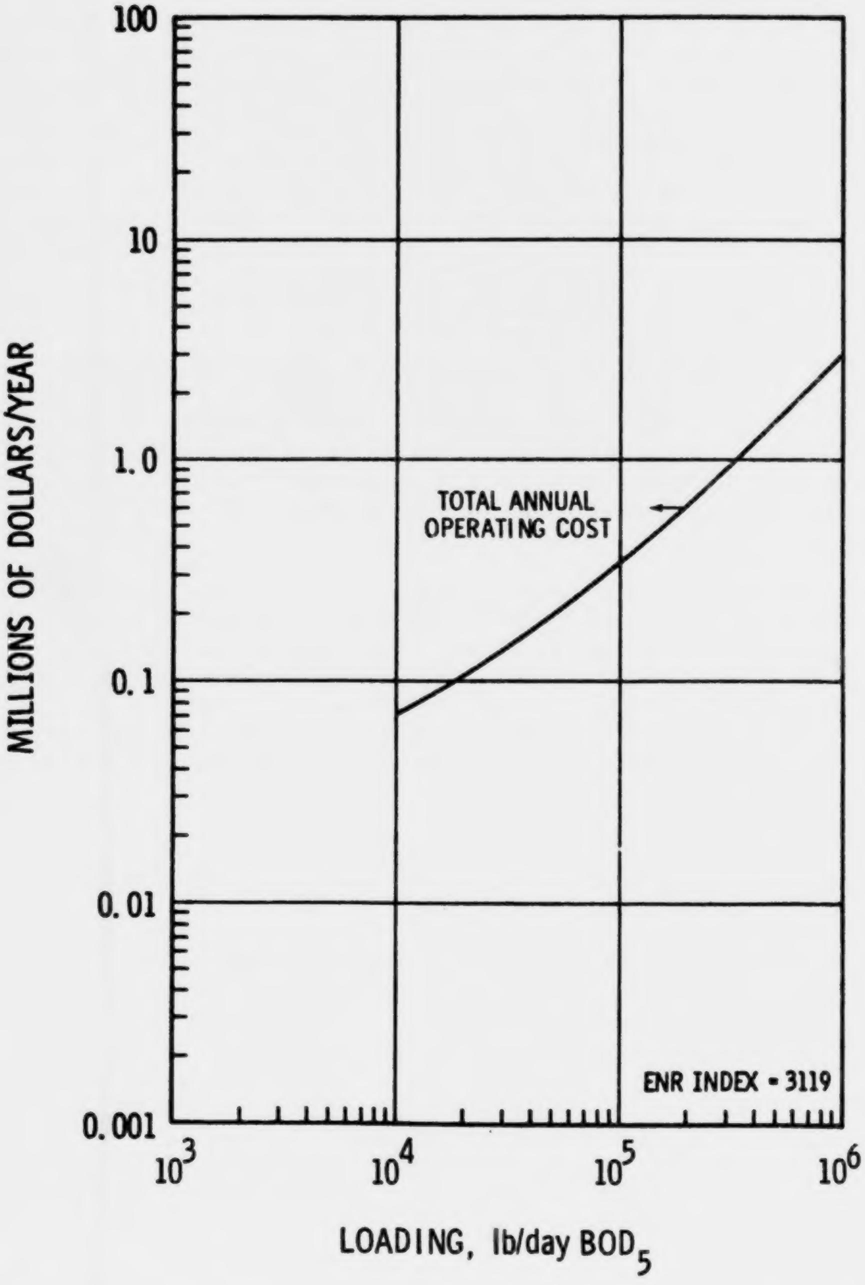
ANNUAL OPERATING COST



IV.4.1.4 Pure Oxygen Activated Sludge, Covered [1]

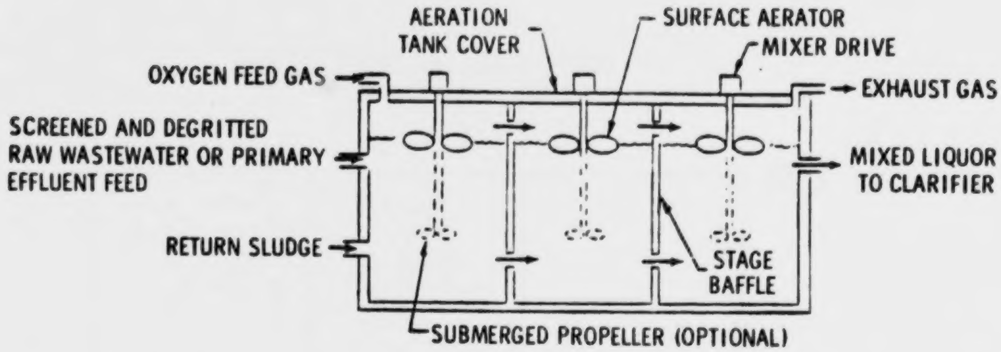
Description. The use of pure oxygen for activated sludge treatment has become competitive with the use of air due to the development of efficient oxygen dissolution systems. The covered oxygen system may be a high-rate activated sludge system. The main benefits cited for the process include reduced power requirements for dissolving oxygen in the wastewater, reduced aeration tank volume requirements, and improved biokinetics of the activated sludge system. In the covered system, oxygenation is performed in a staged, covered reactor in which oxygen gas is recirculated within the system until it reaches a reduced level of purity and a decreased undissolved mass at which it can no longer be used and is vented to the atmosphere. High-purity oxygen gas (90 to 100 percent volume) either from direct on-site generation, off-site generation combined with pipeline delivery, or trucked-in and on-site stored liquid oxygen followed by vaporization enters the first stage of the system and flows concurrently with the wastewater being treated through the aeration basin. Pressure under the tank covers is essentially atmospheric, being held at 2 to 4 inches water column, sufficient to maintain oxygen gas feed control and prevent backmixing from stage to stage. Effluent mixed liquor is separated in conventional gravity clarifiers, and the thickened sludge is recycled to the first stage for contact with influent wastewater.

Mass transfer and mixing within each stage are accomplished either with surface aerators or with a submerged-turbine rotating-sparge system. In the first case, mass transfer occurs in the gas phase; in the latter, oxygen is sparged into the mixed liquor where mass transfer occurs from the oxygen bubbles to the bulk liquid. In both cases, the mass-transfer process is enhanced by the high oxygen partial pressure maintained under the tank covers in each stage.

The UNOX® and OASES® processes are examples of patented and licensed systems, respectively, for pure oxygen activated sludge based on the description presented here.

Although flexibility is claimed to permit operation in any of the normally used flow regimes, i.e., plug flow, complete mix, step aeration, and contact stabilization, the method of oxygen contact employed favors the plug-flow mode.

Equipment normally associated with covered pure-oxygen activated sludge systems include the aeration basin, oxygen generators, liquid oxygen storage tank (for standby and peak load capacity), and aerators. A diagram of the covered pure-oxygen activated sludge process is shown on the next page.


AERATION TANK COVER
SURFACE AERATOR
MIXER DRIVE
OXYGEN FEED GAS
EXHAUST GAS
SCREENED AND DEGRITTED RAW WASTEWATER OR PRIMARY EFFLUENT FEED
MIXED LIQUOR TO CLARIFIER
RETURN SLUDGE
STAGE BAFFLE
SUBMERGED PROPELLER (OPTIONAL)

Typical design criteria. Typical design criteria for a covered pure-oxygen activated sludge system are listed below.

Design criteria	Value
Food-to-microorganism (F/M) ratio	0.5 - 1.0 lb BOD_5/d/lb MLVSS
Mixed liquor suspended solids (MLSS)	3,000 - 6,000 mg/L
Mixed liquor dissolved oxygen	4 - 8 mg/L
Oxygen required	1.2 lb O_2/lb BOD_5 removed

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes oxygenation basins and covers, dissolution equipment, oxygen generators and liquid oxygen feed, storage facilities, instrumentation, and licensing fees. The operating characteristics shown in the table below were assumed for cost estimation on the assumptions that (1) oxygen is delivered as liquid oxygen for plants ranging in size from 0.1 to 1 Mgpd, and (2) oxygen is generated on site for plants ranging in size from 1.0 to 100 Mgpd.

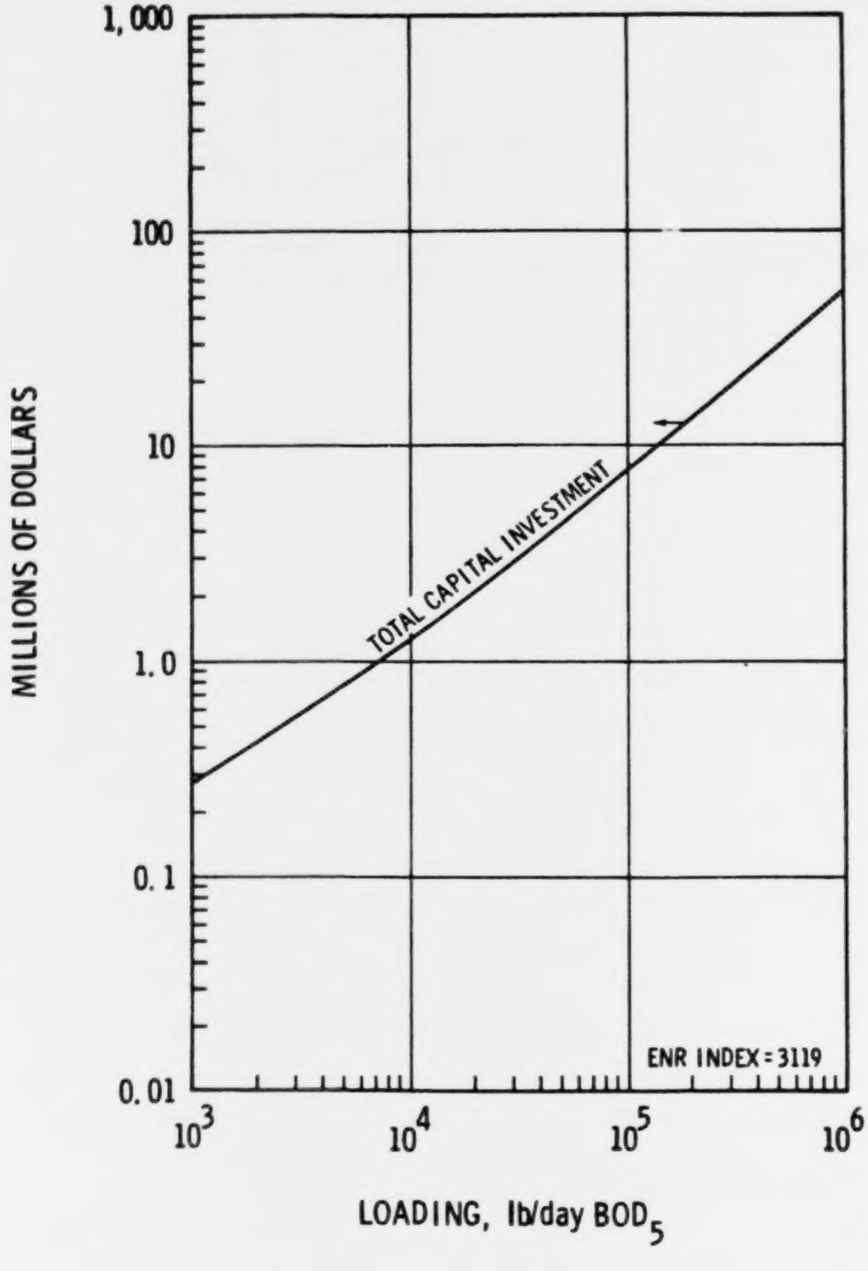
Operating characteristic	Assumed value
Organic loading	1,200 mg/L BOD_5
Oxygen supplied	1.2 lb O_2/lb BOD_5 removed
MLVSS[a]	3,000 mg/L
F/M	0.5 lb BOD_5/d/lb MLVSS

[a] Mixed liquor volatile suspended solids.

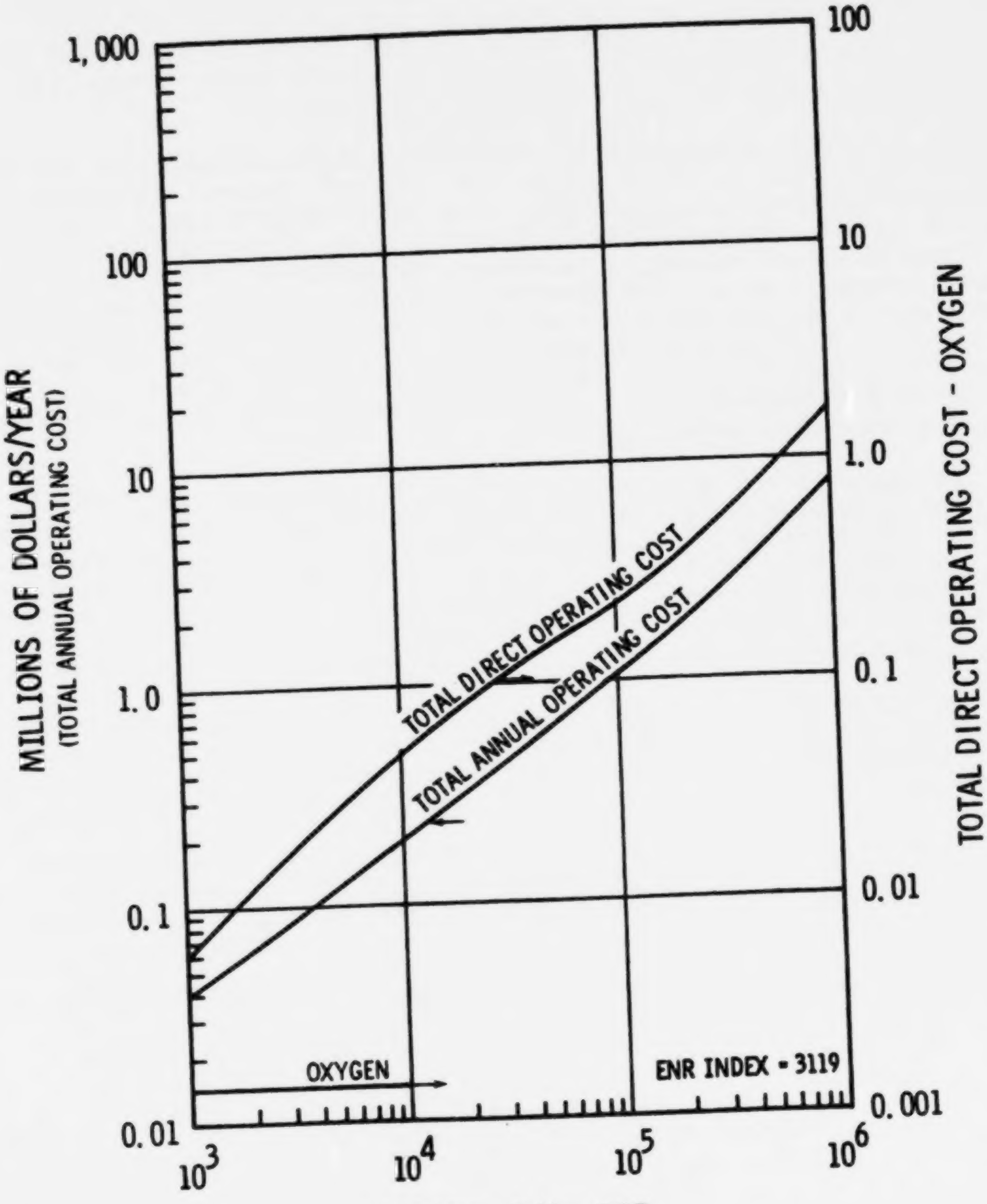
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



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IV.4.1.5 Pure-Oxygen Activated Sludge (Uncovered) [1]

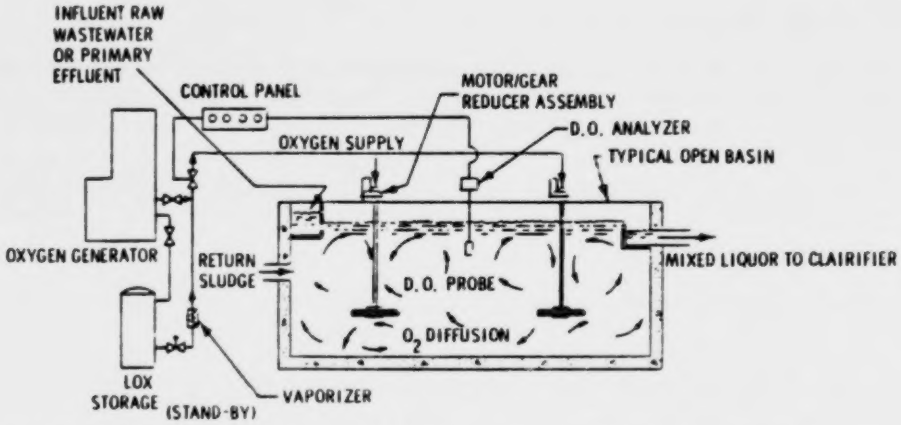
Description. In the uncovered system, oxygenation is performed in an open reactor in which extremely fine porous diffusers are utilized to develop small oxygen gas bubbles that are completely dissolved before breaking surface in normal-depth tanks. The principles that apply in the transfer of oxygen in conventional diffused air systems also apply to the open-tank, pure-oxygen system.

The pure-oxygen, open-tank system currently available is the FMC system (formerly referred to as the "Marox" system) in which ultrafine bubbles are produced, with a correspondingly high gas-surface area. These ultrafine bubbles are of micron size, whereas "fine bubbles" normally produced in diffused air systems are in millimeter sizes. The complete oxygenation system is composed of an oxygen dissolution system comprised of rotating diffusers; a source of high-purity oxygen gas (normally, an on-site oxygen generator); and an oxygen control system, which balances oxygen supply with oxygen demand through use of basin-located dissolved-oxygen probes and control valves.

The influent to the system enters the oxygenation tank and is mixed with return activated sludge. The mixed liquor is continuously and thoroughly mixed using low-energy mechanical agitation near the bottom of the tank. Mixing is produced by radial turbine impellers located on both surfaces (top and bottom) of the rotating diffusion discs. Pure-oxygen gas in the form of micron-size bubbles is simultaneously introduced into the tank to accomplish oxygen transfer. The rotating diffuser is a gear-driven disc-shaped device equipped with a porous medium to assist in the diffusion process. As the diffuser rotates at constant speed in the mixed liquor, hydraulic shear wipes bubbles from the medium before they have an opportunity to coalesce and enlarge.

Operation in any of the normally-used flow regimes (i.e., plug flow, complete mix, step aeration and contact stabilization) can be used as conditions dictate because the method of oxygen contact employed does not favor one particular operating mode.

A flow diagram of the uncovered pure-oxygen activated sludge process is shown on the next page.


INFLUENT RAW WASTEWATER OR PRIMARY EFFLUENT
CONTROL PANEL
MOTOR/GEAR REDUCER ASSEMBLY
D.O. ANALYZER
OXYGEN SUPPLY
TYPICAL OPEN BASIN
OXYGEN GENERATOR
RETURN SLUDGE
MIXED LIQUOR TO CLAIRIFIER
D.O. PROBE
O2 DIFFUSION
LOX STORAGE
(STAND-BY)
VAPORIZER

Typical design criteria. Typical design criteria for the uncovered pure-oxygen activated sludge system are shown in the following table.

Design criteria	Value
Food-to-microorganism (F/M) ratio	0.5 - 1.0 lb BOD_5/d/lb MLVSS
Mixed liquor dissolved oxygen	2 - 6 mg/L
Mixed liquor suspended solids (MLSS)	3,000 - 6,000 mg/L

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes oxygenation basins, dissolution equipment, oxygen generators and liquid oxygen feed, storage facilities, instrumentation (where applicable) and licensing fees. The operating characteristics shown in the table below were assumed for cost estimation on the assumptions that (1) oxygen is delivered as liquid oxygen for plants with loads up to about 1.1 x 10^3 lb/d BOD, and (2) oxygen is generated on site for plants treating loads greater than 1.1 x 10^3 lb/d BOD.

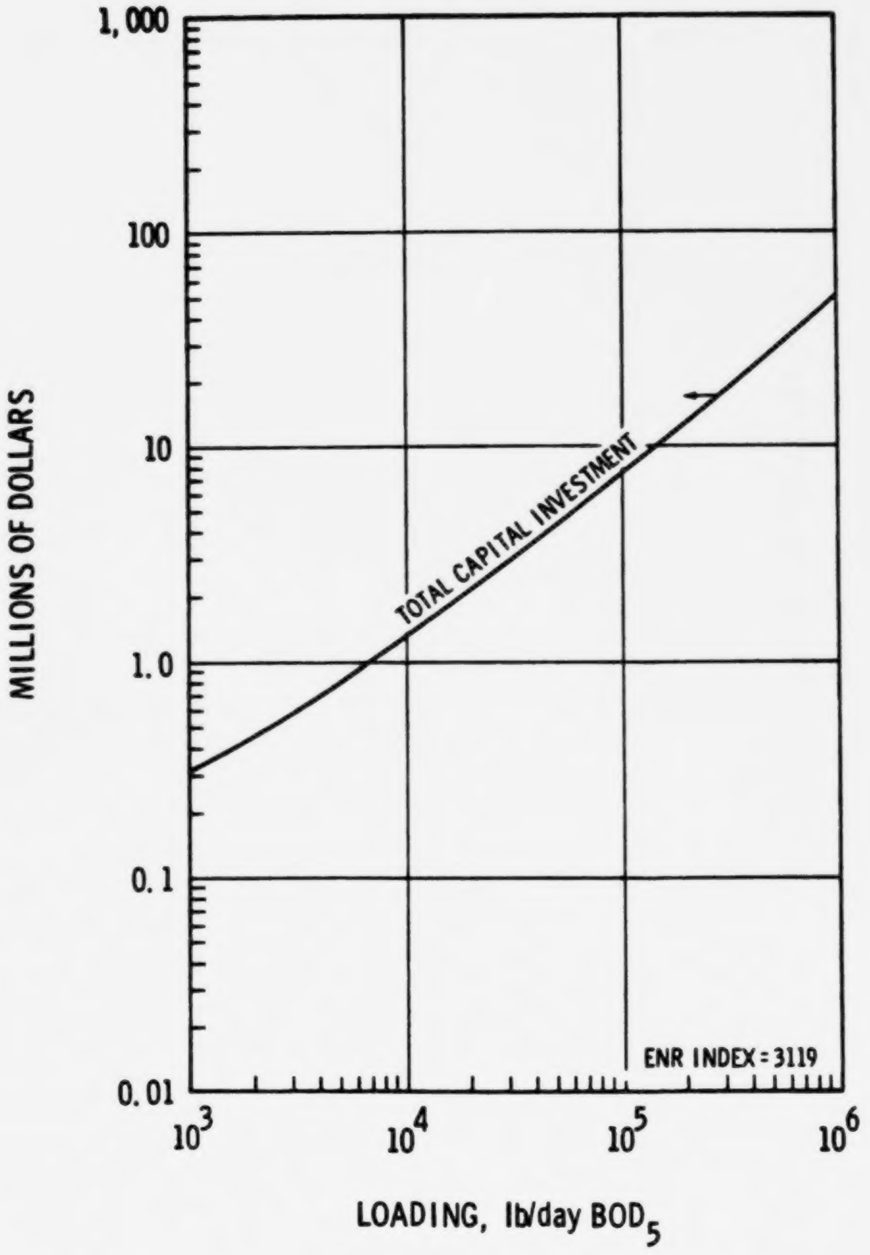
Operating characteristic	Assumed value
Organic loading	1,200 mg/L BOD_5
Oxygen supplied	1.2 lb O_2/lb BOD_5 removed
MLVSS[a]	3,000 mg/L
F/M	0.5 lb BOD_5/d/lb MLVSS

[a] Mixed liquior volatile suspended solids.

References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



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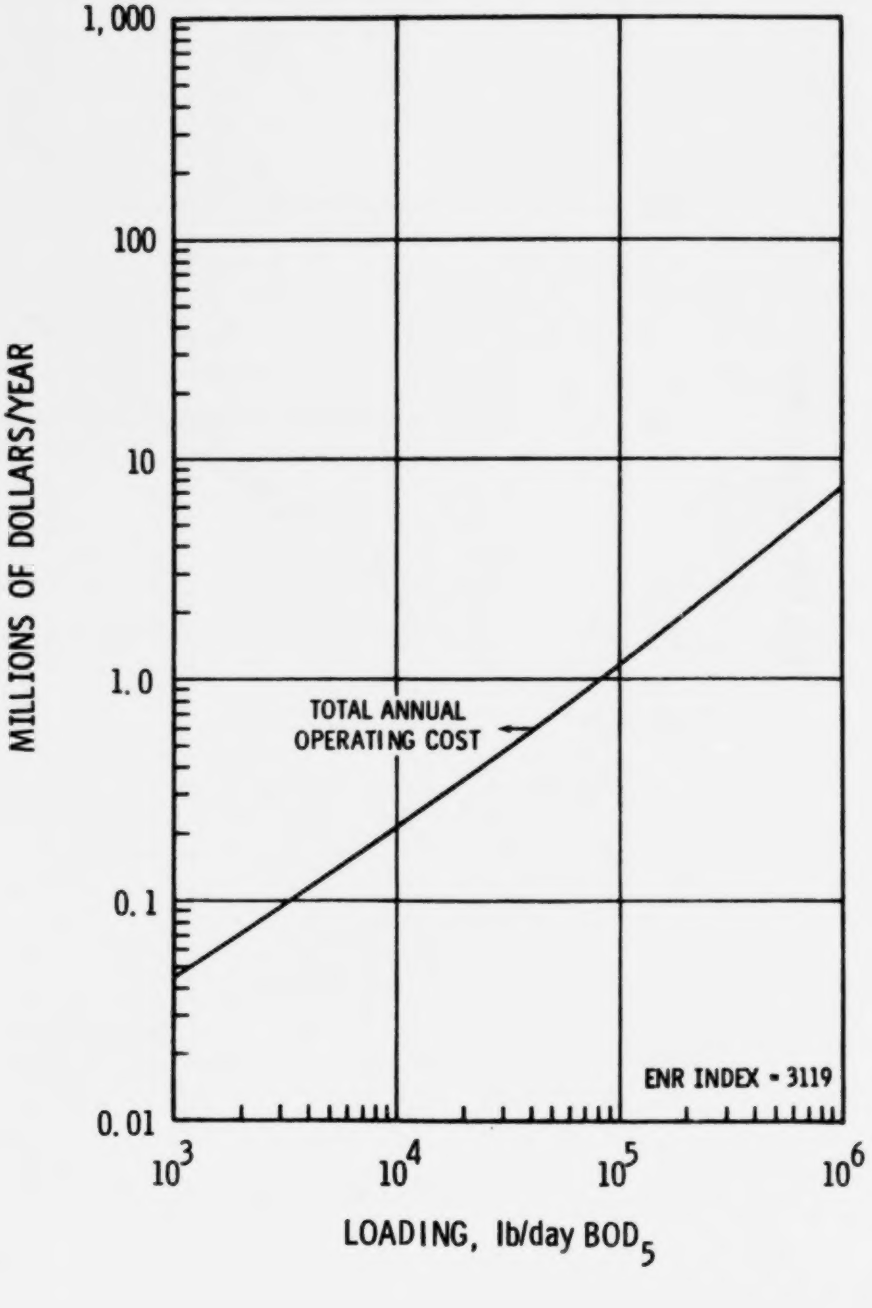
ANNUAL OPERATING COST



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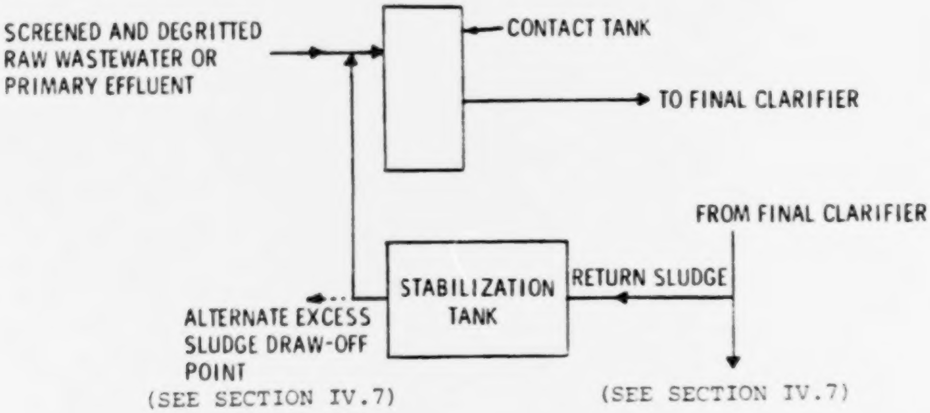
IV.4.1.6 Contact Stabilization, Diffused Aeration [1]

Description. Contact stabilization is a modification of the activated sludge process. In this modification, the adsorptive capacity of the floc is utilized in the contact tank to adsorb suspended, colloidal, and some dissolved organics. The hydraulic detention time in the contact tank is only 30 to 60 minutes (domestic). After the biological sludge is separated from the wastewater in the secondary clarifier, the concentrated sludge is separately aerated in the stabilization tank with a detention time of 2 to 6 hours (domestic). The adsorbed organics undergo oxidation in the stabilization tank and are synthesized into microbial cells. If the detention time is long enough in the stabilization tank, endogenous respiration will occur, along with a concomitant decrease in excess biological sludge production. Following stabilization, the reaerated sludge is mixed with incoming wastewater in the contact tank, and the cycle starts anew.

This process requires smaller total aeration volume than the conventional activated sludge process. It also can handle greater organic shock and toxic loadings because of the biological buffering capacity of the stabilization tank and the fact that at any given time the majority of the activated sludge is isolated from the main stream of the plant flow. Generally, the total aeration basin volume (contact plus stabilization basins) is only 50% to 75% of that required in the conventional activated sludge system.

Equipment normally associated with the contact stabilization process includes air diffusers, compressors, clarifier, contact basin, and stabilization basin. A flow diagram of a contact stabilization system is shown below.


SCREENED AND DEGRITTED RAW WASTEWATER OR PRIMARY EFFLUENT
CONTACT TANK
TO FINAL CLARIFIER
FROM FINAL CLARIFIER
STABILIZATION TANK
RETURN SLUDGE
ALTERNATE EXCESS SLUDGE DRAW-OFF POINT
(SEE SECTION IV.7)
(SEE SECTION IV.7)

Typical design criteria. A partial listing of typical design criteria for the contact stabilization process is summarized below.

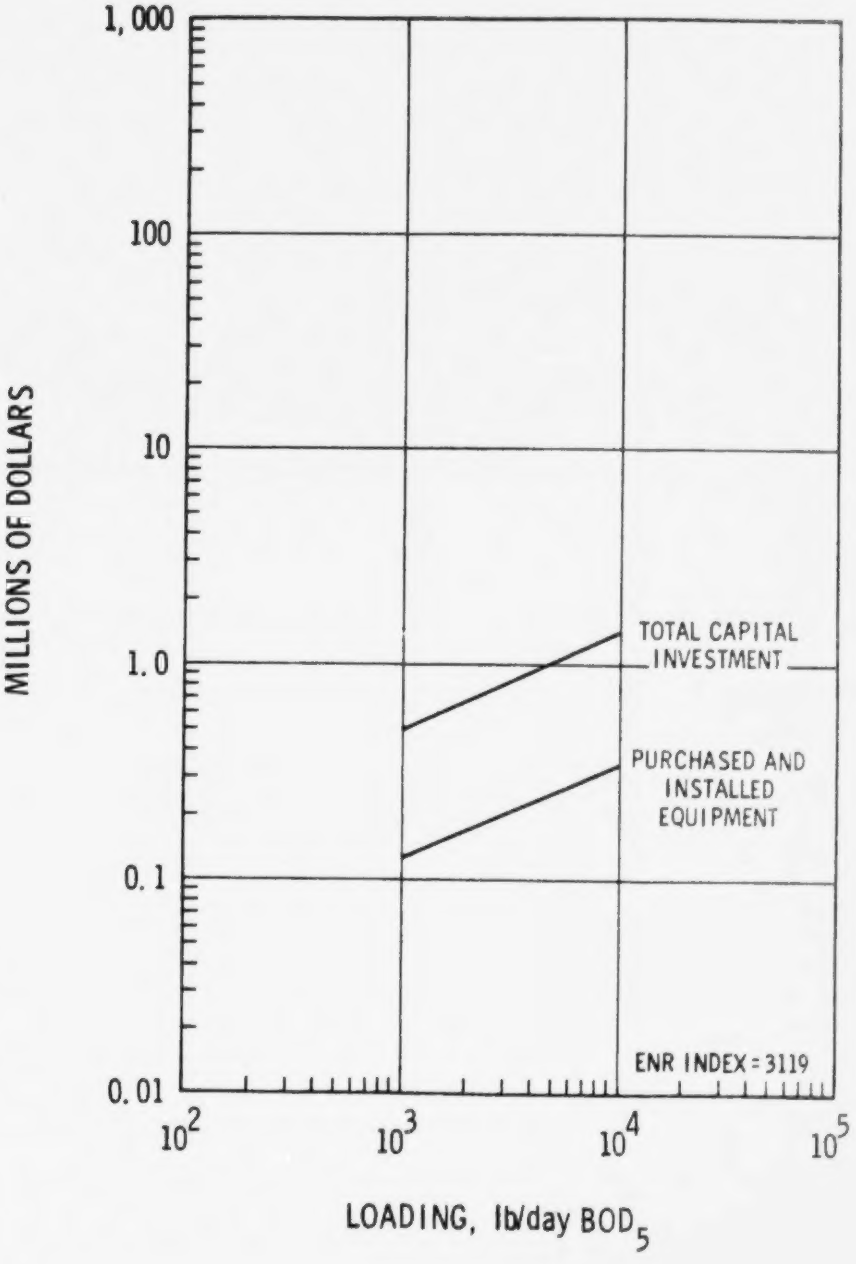
Design criteria	Value
Organic loading	1,200 mg/L BOD_5
Food-to-microorganism (F/M) ratio	0.2 - 0.6 lb BOD_5/d/lb MLSS
Mixed liquor suspended solids (MLSS)	1,000 - 2,500 mg/L for contact tank; 4,000 - 10,000 mg/L for stabilization tank.
Sludge retention time	5 - 10 d
Recycle ratio	0.25 - 1.0
Oxygen required	0.7 - 1.0 lb O_2/lb BOD_5 removed
Volatile fraction of MLSS	0.6 - 0.8

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes aeration basin, clarifier, chlorine contact chamber, aerobic digester, and chlorine feed facility. An organic loading of 1,200 mg/L BOD_5 was used.

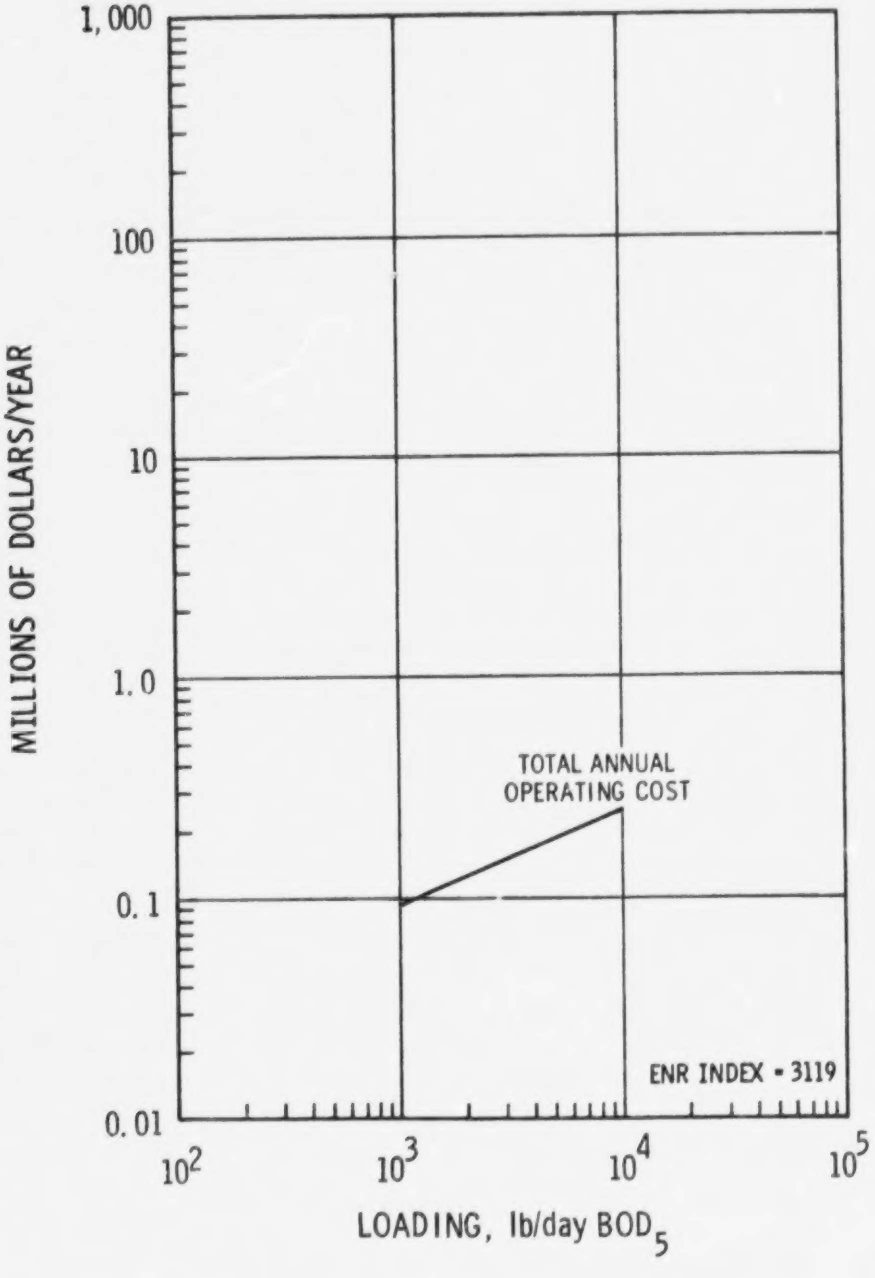
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



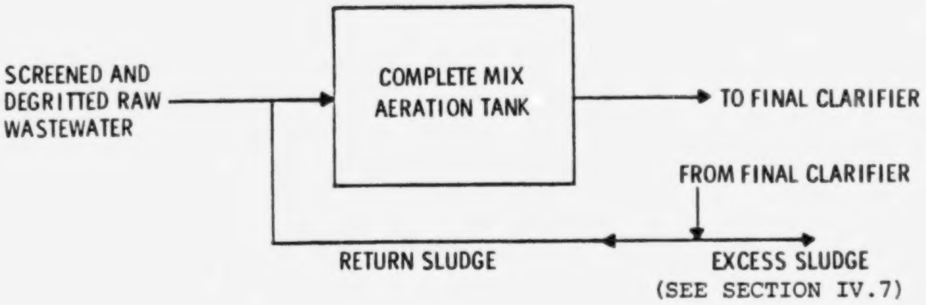
IV.4.1.7 Extended Aeration, Mechanical and Diffused Aeration [1]

Description. Extended aeration is the "low-rate" modification of the activated sludge process. The food-to-microorganism (F/M) loading is in the range of 0.05 to 0.15 lb BOD_5/d/lb MLVSS. The extended aeration system operates in the endogenous respiration phase of the bacterial growth cycle, because of the low BOD_5 loading. The organisms are starved and forced to undergo partial auto-oxidation.

In the complete mix version of the extended aeration process, all portions of the aeration basin are essentially homogeneous, resulting in a uniform oxygen demand throughout the aeration tank. This condition can be accomplished fairly simply in a symmetrical (square or circular) basin with a single mechanical aerator or by diffused aeration. The raw wastewater and return sludge enter at a point (e.g., under a mechanical aerator) where they are quickly dispersed throughout the basin. In rectangular basins with mechanical aerators or diffused air, the incoming waste and return sludge may be distributed along one side of the basin, and the mixed liquor is withdrawn from the opposite side.

Common modifications include step aeration, contact stabilization, and plug flow regimes.

Equipment normally associated with the extended aeration process includes aerators, aeration tanks, clarifier, air diffusers, and compressors. A flow diagram of an extended aeration system is shown below.



Typical design criteria. A partial listing of typical design criteria for the extended aeration activated sludge process is summarized below.

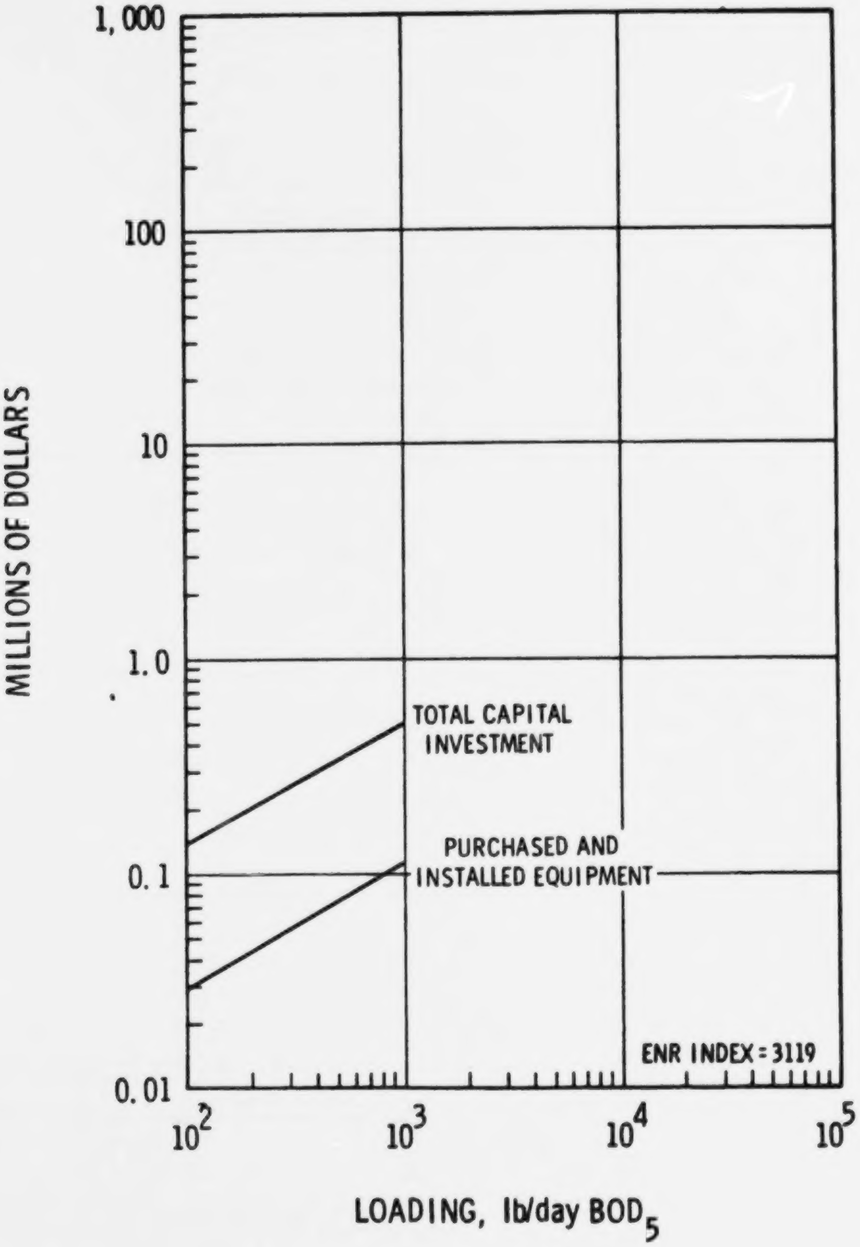
Design criteria	Value
Mixed liquor suspended solids (MLSS)	3,000-6,000 mg/L
Food-to-microorganism (F/M) ratio	0.05-0.15 lb BOD_5/d/lb MLVSS
Oxygen required	2.0-2.5 lb O_2/lb BOD_5 on 1.5 lb O_2/lb BOD_5 removed and 4.6 lb O_2/lb NH_4-N removed)
Sludge age	20-40 d
Recycle ratio	0.75-1.5
Volatile fraction of MLSS	0.6-0.7

Costs. Construction cost includes comminutor, aeration basin, clarifier, chlorine contact chamber, aerobic digester, and chlorine feed facility. Detention time: 24 hours (based on average daily flow). An organic loading of 1,200 mg/L BOD_5 was used.

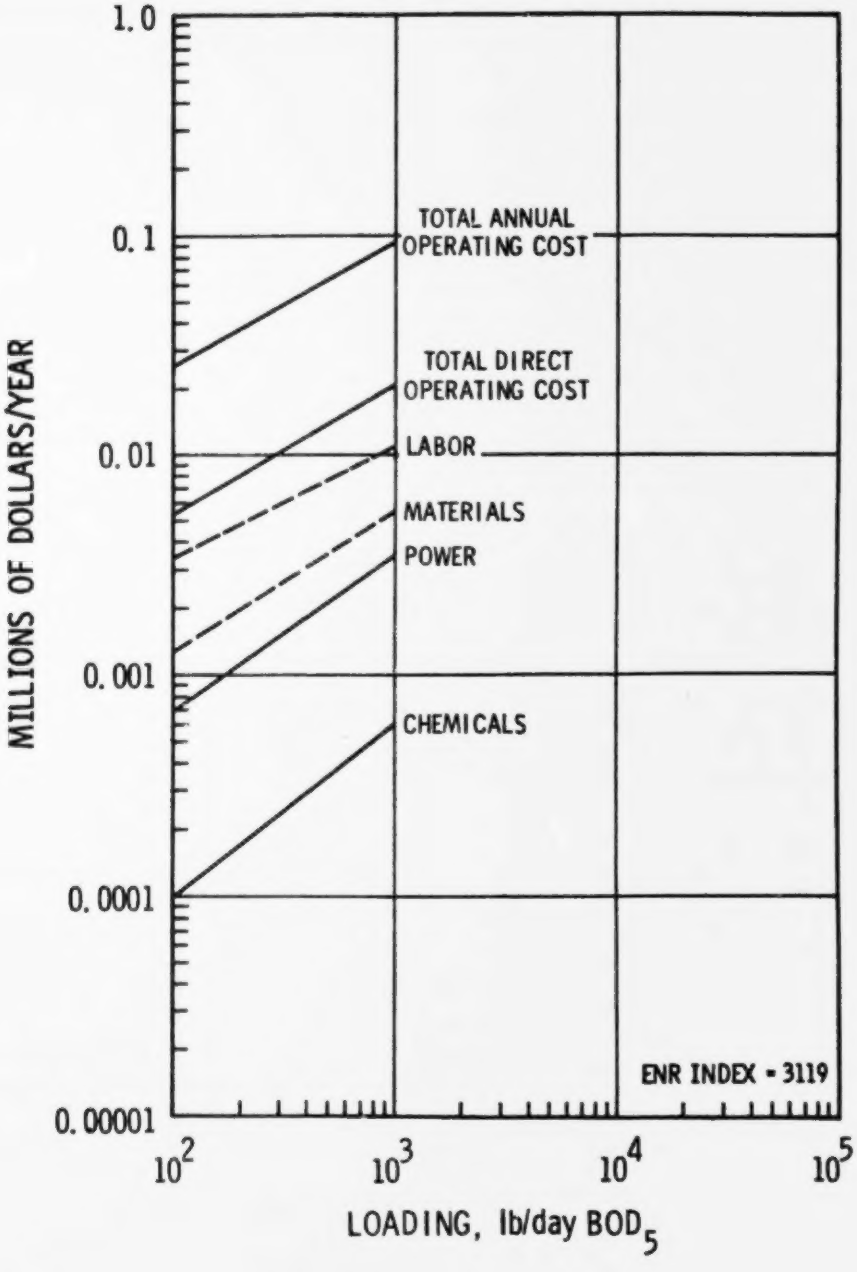
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



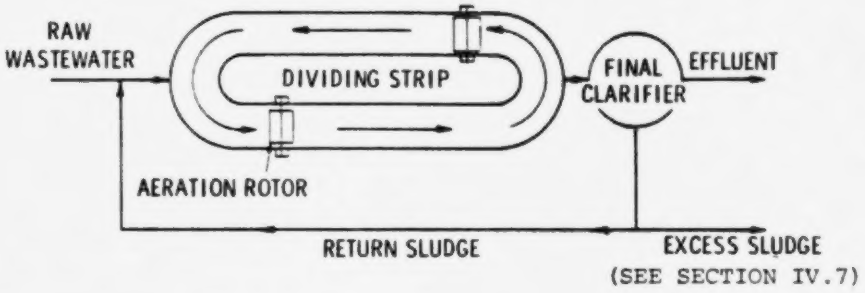
IV.4.1.8 Oxidation Ditch [1]

Description. An oxidation ditch is an activated sludge biological treatment process, which is commonly operated in the extended aeration mode, although conventional activated sludge treatment is also possible. Typical oxidation ditch treatment systems consist of a single or closed loop channel, 4 to 6 feet deep, with 45° sloping sidewalls.

Some form of preliminary treatment such as screening, comminution or grit removal may precede the process After pretreatment the wastewater is aerated in the ditch using mechanical aerators that are mounted across the channel. Horizontal brush, cage, or disc-type aerators, as well as jet aerators, specially designed for oxidation ditch applications, are normally used. The aerators provide mixing and circulation in the ditch, as well as sufficient oxygen transfer. Mixing in the channels is uniform, but zones of low dissolved oxygen concentration can develop. Aerators operate in the 60 to 110 rpm range and provide sufficient velocity to maintain solids in suspension. A high degree of nitrification may occur in the process without special modification when high solid retention times (10 to 50 days) are utilized. Secondary settling of the aeration ditch effluent is provided in a separate clarifier.

Ditches may be constructed of various materials, including concrete, gunite, asphalt, or impervious membranes; concrete is the most common. Ditch loops may be oval or circular in shape. "Ell" and "horseshoe" configurations have been constructed to maximize land usage. Conventional activated sludge treatment, in contrast to extended aeration, may be practiced. Oxidation ditch systems with depths of 10 feet or more with vertical sidewalls and vertical shaft aerators may also be used.

Equipment normally associated with an oxidation ditch system includes hydraulic controls and brush aerators. A flow diagram of the oxidation ditch process is shown below.



Typical design criteria. Typical design criteria for an oxidation ditch system operated in the extended aeration mode are listed below.

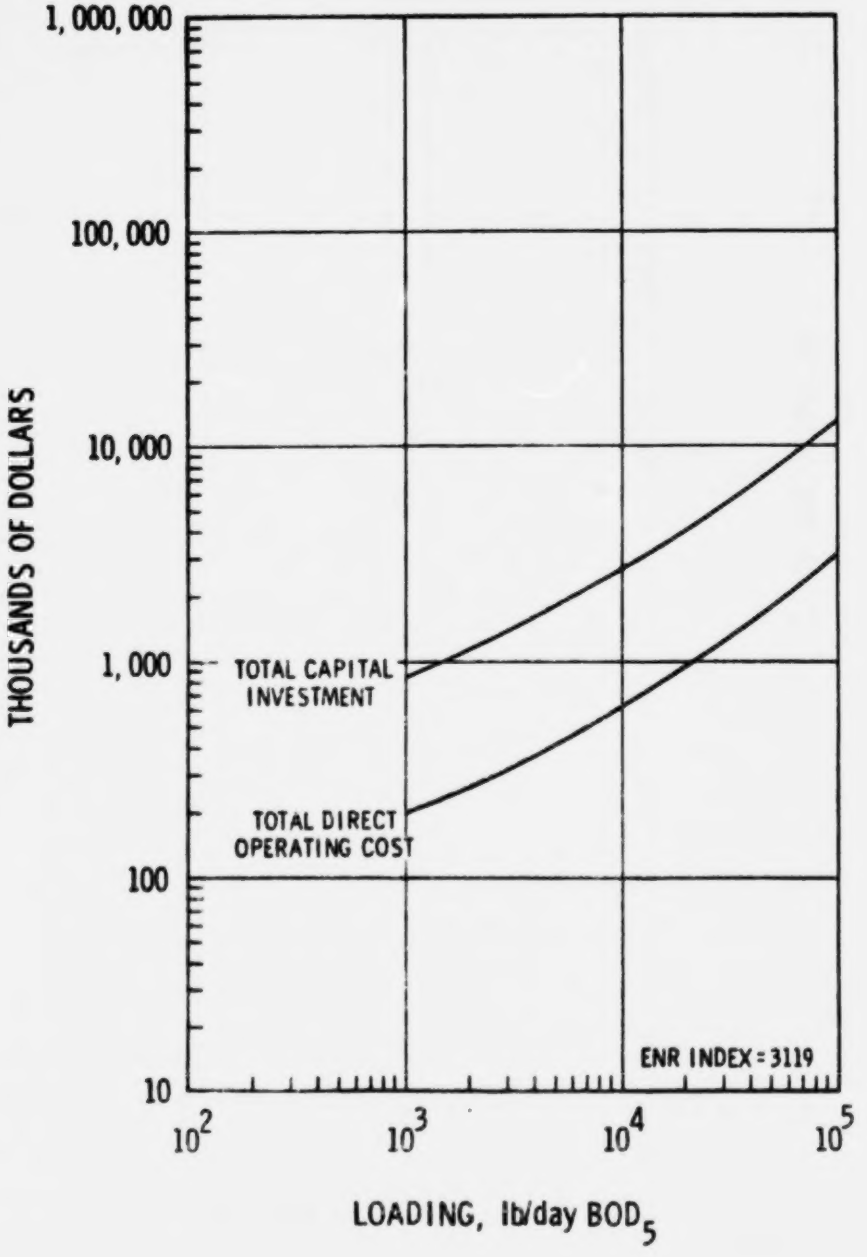
Design criteria	Value
Organic loading	1,200 mg/L BOD_5
Sludge age	10-50 d
Channel depth	4-6 ft
Channel geometry	45° or vertical sidewalls
Aeration channel detention time	Depends on waste strength

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes building, laboratory, and outdoor sludge drying beds.

References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



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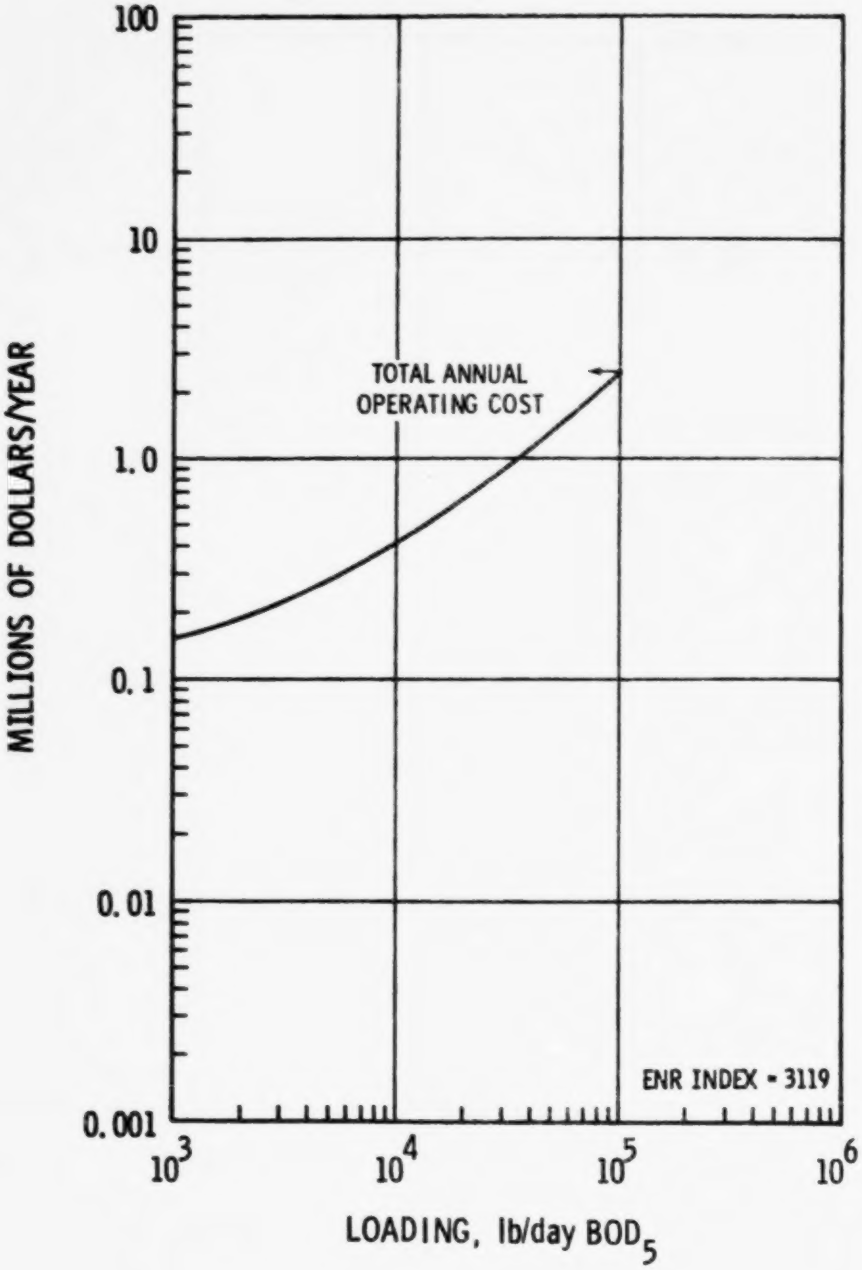


IV.4.2 TRICKLING FILTERS

Trickling filters are applicable to the same industrial waste streams as the activated sludge treatment process. In the trickling filter process, wastes are sprayed through the air to absorb oxygen and allowed to trickle through a bed of rock or synthetic media coated with a slime of microbial growth. The microbial slime is capable of decomposing organic matter in the waste stream. Process modifications employ various types and depths of media to retain the microorganisms under varying hydraulic conditions [1].

Trickling filters provide media for the support of biomass, rather than have suspended biomass as do activated sludge treatment processes. The wastewater is trickled through the media and collected underneath in a drain. Suspended solids, including microbial slime that has sloughed off the media, are clarified from the underflow, which may be either recycled to the trickling filter head or sent to following treatment units. Hydraulic loading rates for trickling filters are generally less than 0.5 gpm/ft^2; however, loadings up to 4 gpm/ft^2 have been employed. Rock media used to pack the filters typically allow depths of 3 to 8 ft. The use of synthetic media allows deep filters because of their lessened tendency to clog and become anaerobic. Also, the comparative light weight of synthetic media allows higher packings. Filter packings up to 40-ft in depth are reasonable where synthetic media are employed [1].

Trickling filtration is classified as low rate, using rock media. and high rate, using rock, plastic and wood media. Plastic media, high-rate rock media, and low-rate rock media trickling filters are discussed in Sections IV.4.2.1 through IV.4.2.3, respectively.

IV.4.2.1 Plastic Media Trickling Filter [2]

Description. The process consists of a fixed bed of plastic media over which wastewater is applied for aerobic biological treatment. Zoogleal slimes form on the media which assimilate and oxidize substances in the wastewater. The bed is dosed by a distributor system, and the treated wastewater is collected by an underdrain system. Primary treatment is normally required to optimize trickling filter performance, whereas post-treatment is generally not required to meet secondary standards.

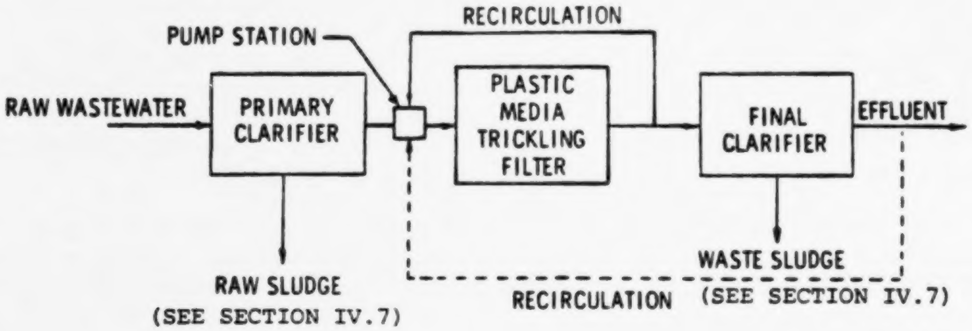
The rotary distributor has become the standard because of its reliability and ease of maintenance, however, fixed nozzles are often used in roughing filters. Plastic media is comparatively light with a specific weight of 10 to 30 times less than rock media. Its high void space (approximately 95 percent) promotes better oxygen transfer during passage of the liquid through the filter than rock media with its approximate 50 percent void space. Because of its light weight, plastic media containment structures are normally constructed as elevated towers 20 to 30 feet high. Excavated containment structures for rock media can sometimes serve as a foundation for elevated towers for converting an existing facility to plastic media.

Plastic media trickling filters can be employed to provide independent secondary treatment or roughing ahead of a second-stage biological process. When used for secondary treatment, the media bed is generally circular in plan and dosed by a rotary distributor. Roughing applications often utilize rectangular media beds with fixed nozzles for distribution.

The organic material present in the wastewater is degraded by a population of microorganisms attached to the filter media. As the microorganisms grow, the thickness of the slime layer increases. Periodically, the liquid will wash some slime off the media, and a new slime layer will start to grow. This phenomenon of losing the slime layer is called sloughing and is primarily a function of the organic and hydraulic loadings on the filter. Filter effluent recirculation is vital with plastic media trickling filters to ensure proper wetting of the media and to promote effective sloughing control compatible with the high organic loadings employed.

Modifications common to all types of trickling filtration include addition of recirculation, multistaging, electrically powered distributors, forced ventilation, filter covers, and use of various methods of pretreatment and post-treatment of wastewater.

Equipment normally associated with plastic media trickling filters includes underdrains, distributors, filter covers, and plastic media. A flow diagram for a plastic media trickling filtration system is shown below.



Typical design criteria. Typical design criteria for plastic media trickling filters include hydraulic loading and organic loading as noted below.

Design criteria	Value
Hydraulic loading (with recirculation)	
Secondary treatment	30-60 Mgal/acre/d; 700-1,400 gpd/ft^2
Roughing	100-200 Mgal/acre/d; 2,300-4,600 gpd/ft^2
Recirculation ratio	0.5:1 to 5:1
Dosing interval	Not more than 15 s (continuous)
Sloughing	Continuous
Organic loading	
Secondary treatment	450-2,200 lb BOD$_5$/d/acre ft; 10-50 lb BOD$_5$/d/1,000 ft^3
Roughing	4,500-22,000 lb BOD$_5$/d/acre ft; 100-500 lb BOD$_5$/d/1,000 ft^3
Bed depth	20-30 ft
Power requirement	20-200 hp/Mgal
Underdrain minimum slope	1%

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes plastic media, underdrains, distributors, and tower containment structures. Clarifiers and recirculation equipment are not included in the purchased and installed equipment cost. Annual operating costs include labor and materials; energy costs are not included. The following operating characteristics were assumed for cost estimation:

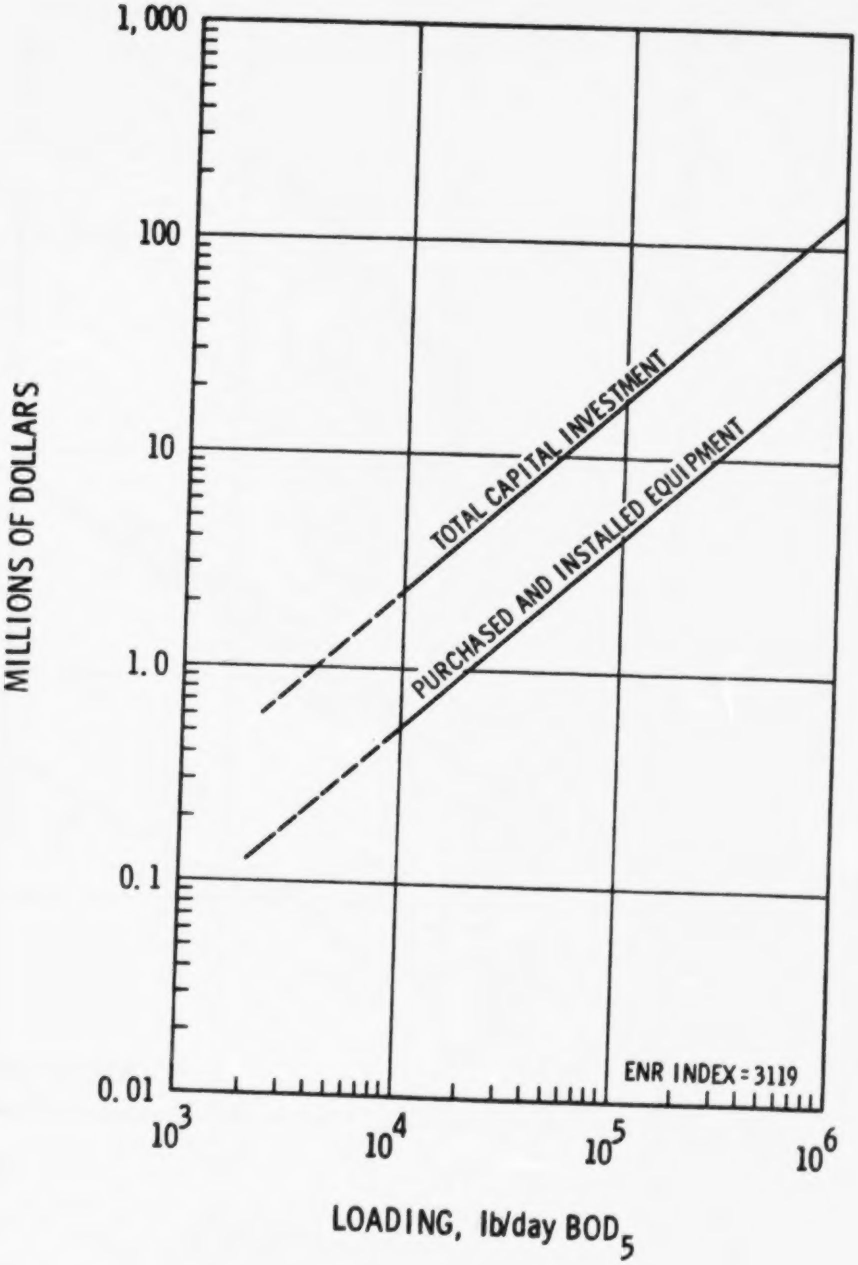
Operating characteristics	Assumed value
Organic loading	1,200 mg/L BOD$_5$
Bed depth	21 ft
BOD$_5$ loading	11 lb BOD$_5$/d/1,000 ft^3
Recirculation rates	4.5 (based on average daily flow) to 0.6 (based on peak daily flow, assumed to be 3.5 times daily average flow)

References.

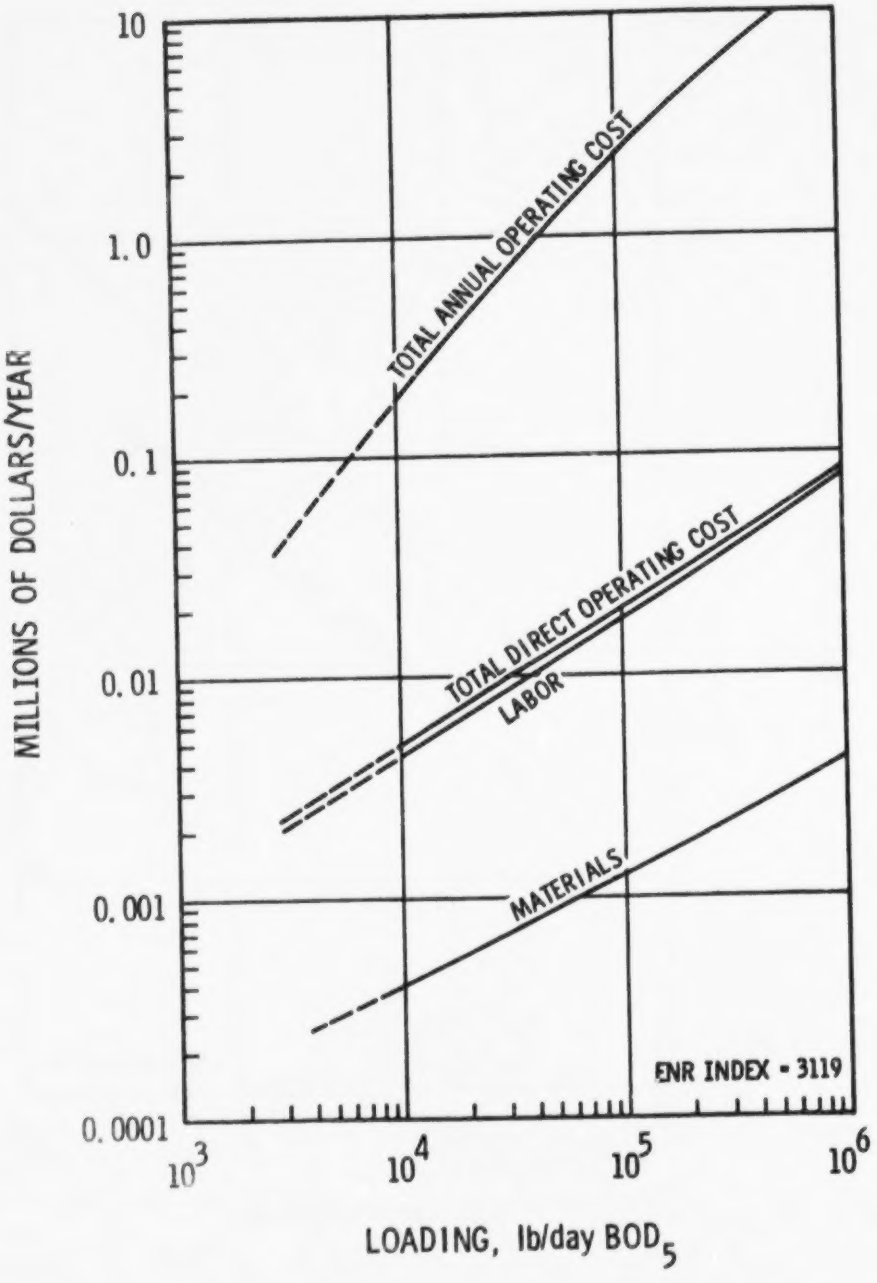
1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 41-1 through 41-15.

2. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.4.2.2 Rock Media Trickling Filter - High Rate [1]

Description. The process consists of a fixed bed of rock media over which wastewater is applied for aerobic biological treatment. Zoogleal slimes form on the media which assimilate and oxidize substances in the wastewater. The bed is dosed by a distributor system, and the treated wastewater is collected by an underdrain system. A primary clarifier is normally used to optimize trickling filter performance, and final treatment is often necessary to meet secondary standards or water quality limitations.

The rotary distributor has become the standard because of its reliability and ease of maintenance. It consists of two or more arms that are mounted on a pivot in the center of the filter. Nozzles distribute the wastewater as the arms rotate due to the dynamic action of the incoming wastewater. Continuous recirculation of filter effluent may be used to maintain a constant hydraulic loading to the distributor arms.

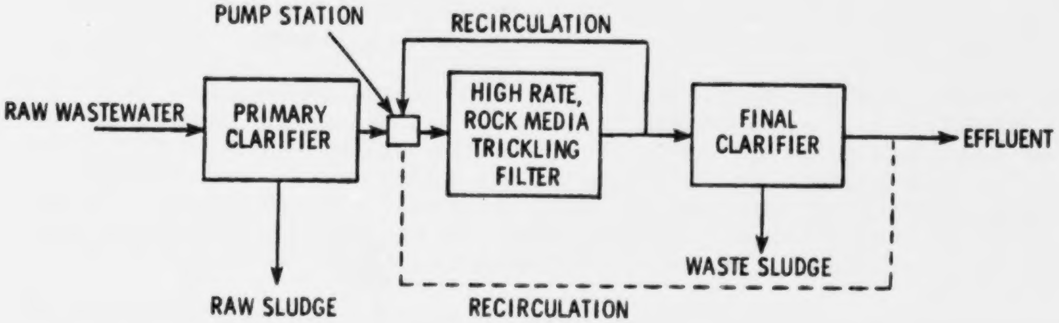
Underdrains are manufactured from specially designed vitrified-clay blocks that support the filter media and pass the filtered solids sewage to a collection sump for transfer to the final clarifier.

The filter media consists of 1- to 5-inch stone. The high rate trickling filter media bed generally is circular in plan, with a depth of 3 to 6 feet. Containment structures are normally made of reinforced concrete and installed in the ground to support the weight of the media.

The organic material present in the wastewater is degraded by a population of microorganisms attached to the filter media. As the microorganisms grow, the thickness of the slime layer increases. As the slime layer increases in thickness, the absorbed organic matter or oxygen is utilized before it can reach the microorganisms near the media face. As a result, the microorganisms near the media face enter into either an endogenous phase or anaerobic phase of growth depending on whether organic matter or oxygen is limiting. In this phase, the microorganims lose the media surface. The liquid then washes the slime off the media, and a new slime layer will start to grow. This phenomenon of losing the slime layer is called sloughing and is primarily a function of the organic and hydraulic loadings on the filter. Filter effluent recirculation is vital with high rate trickling filters to promote the flushing action necessary for effective sloughing control, without which media clogging and anaerobic conditions could develop due to the high organic loading rates employed.

Common modifications include various recirculation methods, rates of recirculation, multistaging, electrically powered distributors, forced ventilation, and filter covers.

Equipment normally associated with high rate, rock media trickling filters includes underdrains and distributors. A flow diagram of a high rate, rock media trickling filtration system is shown below.


PUMP STATION
RECIRCULATION
RAW WASTEWATER
PRIMARY CLARIFIER
HIGH RATE, ROCK MEDIA TRICKLING FILTER
FINAL CLARIFIER
EFFLUENT
WASTE SLUDGE
RAW SLUDGE
RECIRCULATION

Typical design criteria. Typical design criteria for high rate, rock media trickling filters are shown below.

Design criteria	Value
Hydraulic loading (with recirculation)	10-40 Mgal/acre/d; 230-900 gpd/ft^2
Recirculation ratio	0.5:1 to 4:1
Dosing interval	Not more than 15 s (continuous)
Sloughing	Continuous
Media	Rock; 1-5 in.
Organic loading	900-2,600 lb BOD$_5$/d/acre ft; 20-60 lb BOD$_5$/d/1,000 ft^3
Bed depth	3-6 ft
Power requirements	10-50 hp/Mgal
Underdrain minimum slope	1%

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes rock media, underdrains, distributors, and reinforced concrete containment structures. Clarifier and recirculation equipment are not included in the purchased and installed equipment cost. Energy costs are also not included in the annual operating costs. The following operating characteristics were assumed for cost estimation:

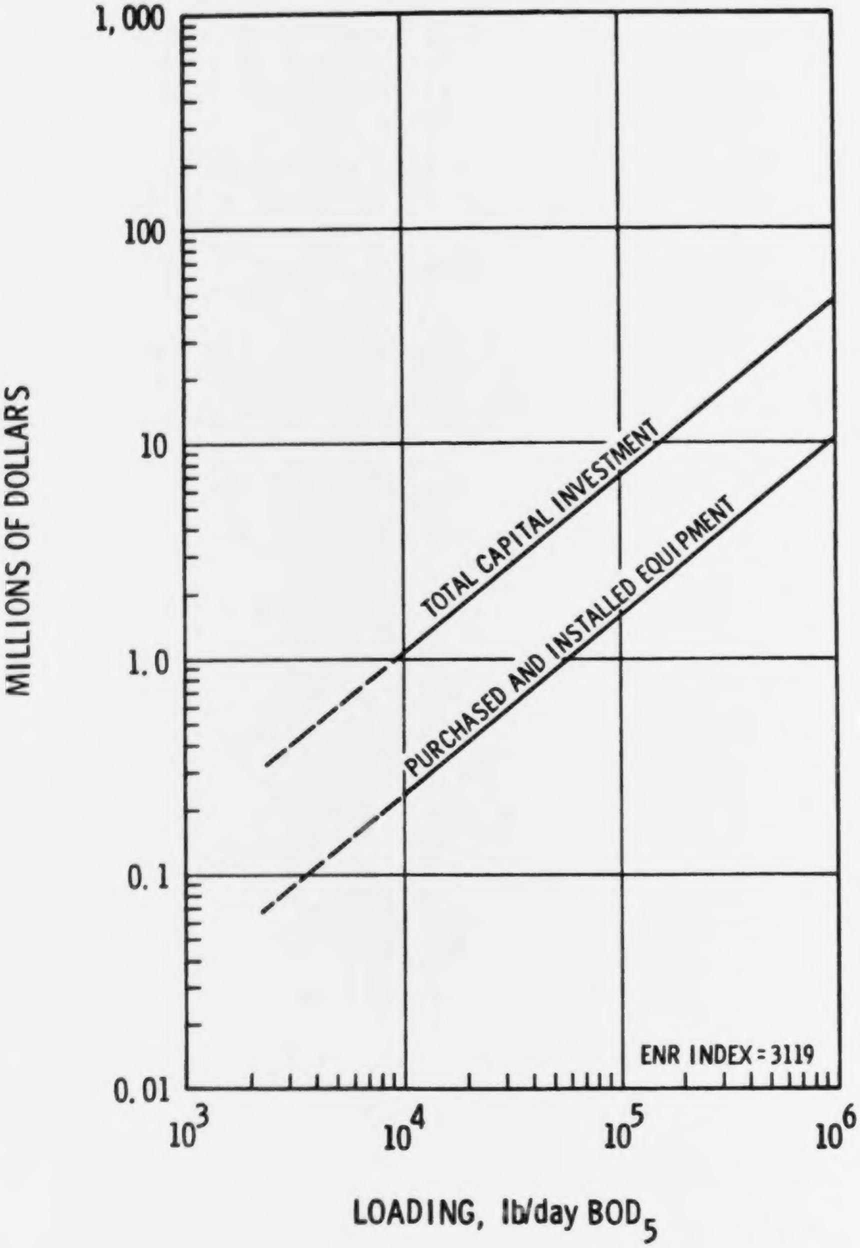
Operating characteristic	Assumed value
Organic loading	1,200 mg/L BOD_5
Bed depth	5 ft
BOD_5 loading	20 lb BOD_5/d/1,000 ft^3 [a]
Hydraulic loading	0.3 gpm/ft^2
Recirculation ratio	4.0 (based on average daily flow) to 0.4 (based on peak daily flow assumed to be 3.5 times average daily flow)

[a] To attain an effluent BOD_5 of 45 mg/L for a domestic sewage influent of 130 mg/L. Based on raw load only.

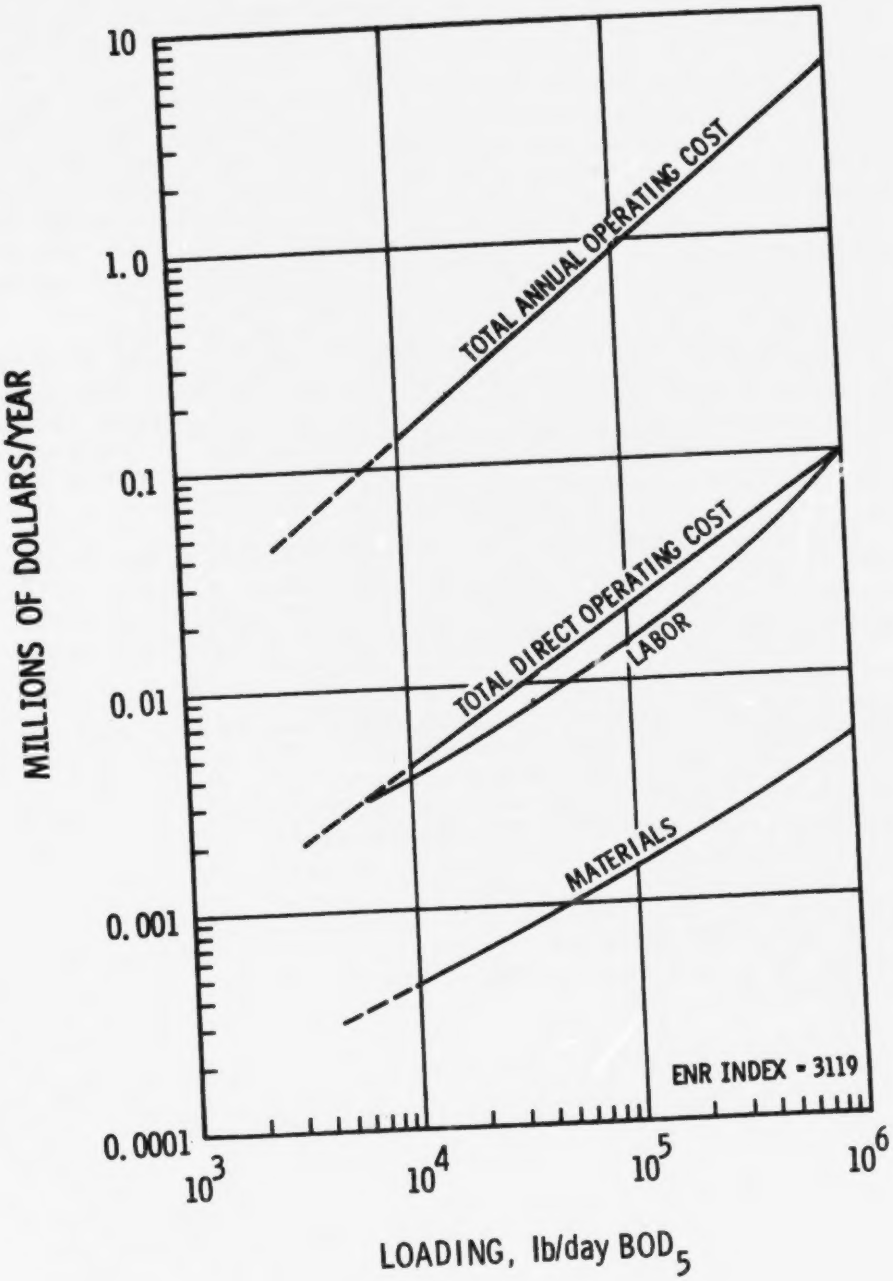
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.4.2.3 Rock Media Trickling Filter - Low Rate [1]

Description. This process also consists of a fixed bed of rock media over which wastewater is applied for aerobic biological treatment. Zoogleal slimes form on the media which assimilate and oxidize substances in the wastewater. The bed is dosed by a distributor system, and the treated wastewater is collected by an underdrain system.

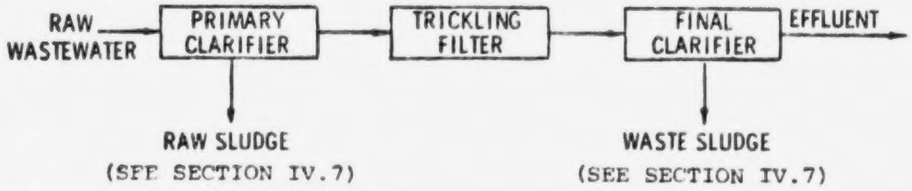
The rotary distributor has become the standard because of its reliability and ease of maintenance. In contrast to the high rate trickling filter which may uses continuous recirculation of filter effluent to maintain a constant hydraulic loading to the distributor arms, either a suction-level controlled pump or a dosing siphon is employed for that purpose with a low rate filter. Nevertheless, programmed rest periods may be necessary at times because of inadequate influent flow.

Underdrains are manufactured from specially designed vitrified-clay blocks that support the filter media and pass the filtered solids to a collection sump for transfer to the final clarifier. The filter media consists of 1- to 5-inch stone. Containment structures are normally made of reinforced concrete and installed in the ground to support the weight of the media.

The low rate trickling filter media bed generally is circular in plan, with a depth of 5 to 10 feet. Although filter effluent recirculation is generally not utilized, it can be provided as a standby tool to keep filter media wet during low flow periods.

The organic material present in the wastewater is degraded by a population of microorganisms attached to the filter media. As the microorganisms grow, the thickness of the slime layer increases. Periodically, wastewater washes the slime off the media, and a new slime layer will start to grow. This phenomenon of losing the slime layer is called sloughing and is primarily a function of the organic and hydraulic loadings on the filter.

Equipment normally associated with low rate, rock media trickling filters includes underdrains, distributors, and filter covers. A flow diagram of a low rate, rock media trickling filtration system is shown below.



Typical design criteria. Typical design criteria for low rate rock media trickling filtration are listed below.

Design criteria	Assumed value
Hydraulic loading	1-4 Mgal/acre/d; 25-90 gpd/ft^2
Organic loading	200-900 lb BOD_5/d/acre ft; 5-20 lb BOD_5/d/1,000 ft^3
Dosing interval	Continuous for majority of daily operating schedule, but may become intermittent (not more than 5 min) during low flow periods
Effluent channel minimum velocity	2 ft/s at average daily flow
Media	Rock; 1-5 in., must meet sodium sulfate soundness test
Depth	5-10 ft
Underdrain minimum slope	1%

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes reinforced concrete containment structures, rotating distributors, rock media, and underdrains. Clarifiers and recirculation equipment are not included in the purchased and installed equipment cost. Energy costs are not included in the annual operating cost. The following operating characteristics were assumed for cost estimation:

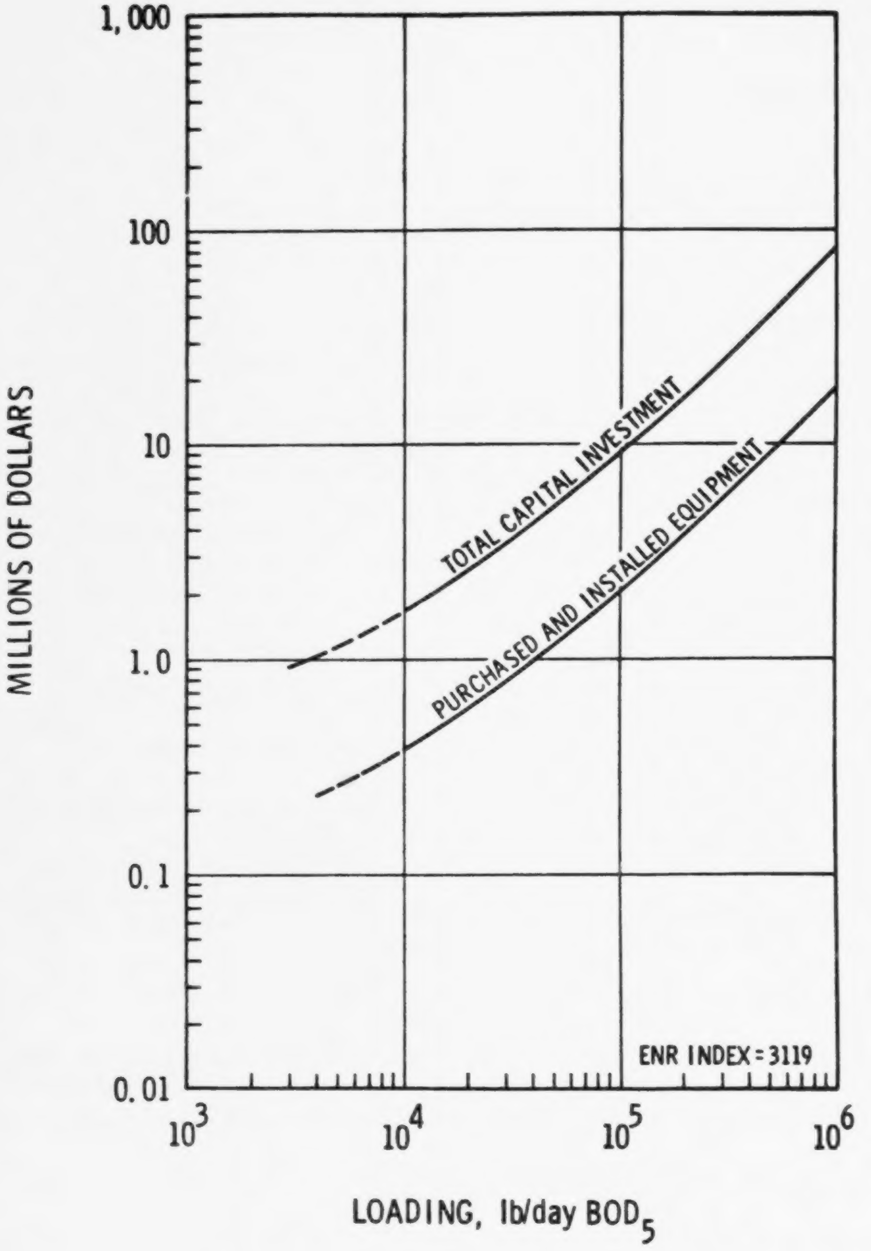
Operating characteristic	Assumed value
Organic layer	1,300 mg/L BOD_5
Bed depth	8 ft
BOD_5 loading	10 lb BOD_5/d/1,000 ft^3[a]
Hydraulic loading	75 gpd/ft^2

[a] To attain an effluent BOD_5 of 25 mg/L at an influent of 1,300 mg/L.

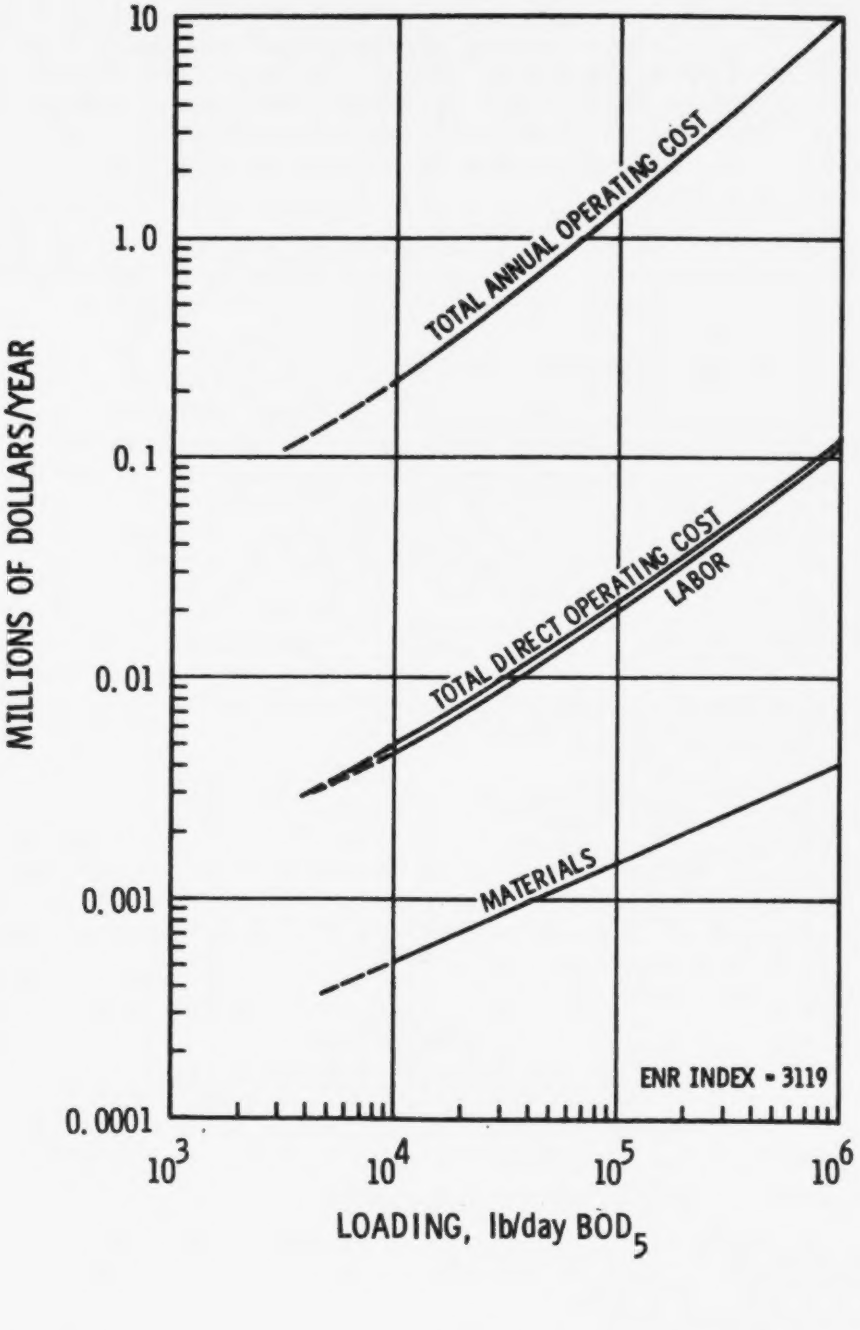
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.4.3 LAGOONS

Lagooning (stabilization ponding) is used to remove dissolved and collodial biodegradable organics, and suspended solids. A lagoon is a relatively shallow body of water contained in an earthen basin of controlled shape, which is designed for the purpose of treating wastewater. The term "oxidation pond," often used, is synonymous. Ponds have become popular because their low construction and operating costs offer a significant financial advantage over other recognized treatment methods. Ponds are used extensively for the treatment of industrial wastes and mixtures of industrial wastes and domestic sewage that are amenable to biological treatment.

Sections 4.3.1 through 4.3.3, respectively, include of discussions on the following modifications of lagoons: aerated lagoons, anaerobic lagoons, and facultative lagoons.

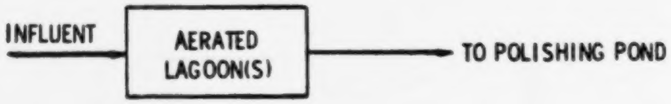
IV.4.3.1 Aerated Lagoons [1]

Description. Aerated lagoons are medium-depth basins designed for the biological treatment of wastewater on a continuous basis. In contrast to stabilization ponds, which obtain oxygen from photosynthesis and surface reaeration, aerated lagoons employ aeration devices that supply supplemental oxygen to the system. The aeration devices may be a mechanical (i.e., surface aerator) or diffused air system. Surface aerators are divided into two types: horizontal aerators, and the more common turbine and vertical shaft aerators. The many diffused air systems utilized in lagoons consist of plastic pipes supported near the bottom of the cells with regularly spaced sparger holes drilled in the tops of the pipes.

Because aerated lagoons are normally designed to achieve partial mixing only, aerobic-anaerobic stratification will occur, and a large fraction of the incoming solids and a large fraction of the biological solids produced from waste conversion settle to the bottom of the lagoon cells. As the solids begin to build up, a portion will undergo anaerobic decomposition. Volatile toxics can potentially be removed by the aeration process, and incidental removal of other toxics can be expected to be similar to an activated sludge system. Several smaller aerated lagoon cells in series are more effective than one large cell. Tapering aeration intensity downward in the direction of flow promotes settling out of solids in the last cell. A nonaerated polishing cell following the last aerated cell is an optional, but recommended, design technique to enhance suspended solids removal prior to discharge.

Lagoons may be lined with concrete or an impervious flexible lining, depending on soil conditions and environmental regulations.

Equipment normally associated with aerated lagoons includes lining systems, aerators, and hydraulic controls. A flow diagram for aerated lagooning is shown below.



Typical design criteria. Typical design criteria for aerated lagoons consist of operation and energy requirements. The operation requires one or more aerated cells, followed by a settling (unaerated) cell. Additional operational criteria are provided below.

Design criteria	Value
Depth	6-20 ft
pH	6.5-8.0
Water temperature range	0-40¼C
Optimum water temperature	20¼C
Oxygen requirement	0.7-1.4 times the amount of BOD_5 removed
Organic loading	10-300 lb BOD_5/acre/d
Operation	One or more cells

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes excavation, embankment, seeding of lagoons/slopes (three cells), service road, fencing rirap embankment protection, hydraulic control works, aeration equipment, and electrical equipment. An organic loading of 2,100 mg/L was assumed for cost estimation.

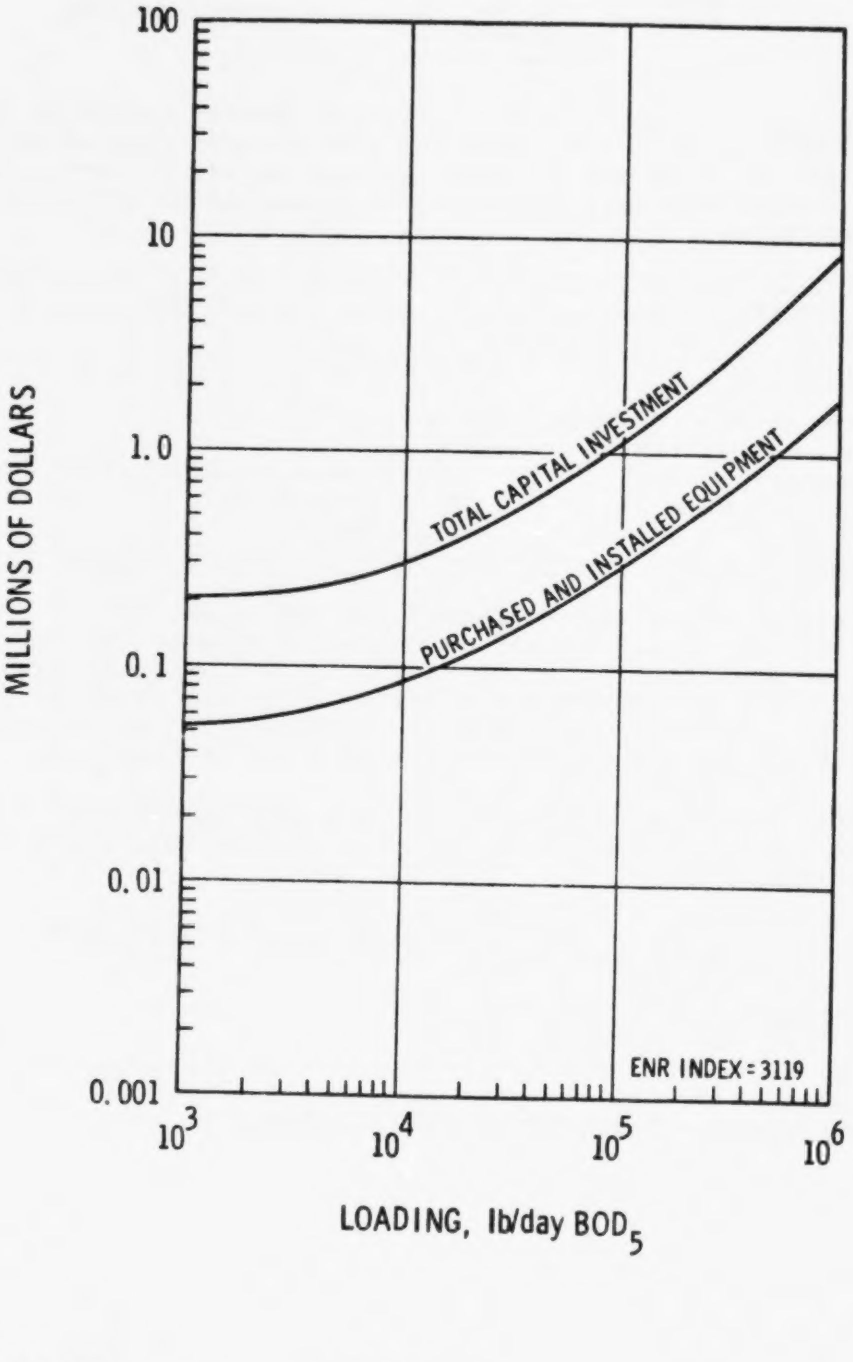
To adjust purchased and installed equipments cost for other detention times, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{\text{new design detection time}}{\text{7 days}}$$

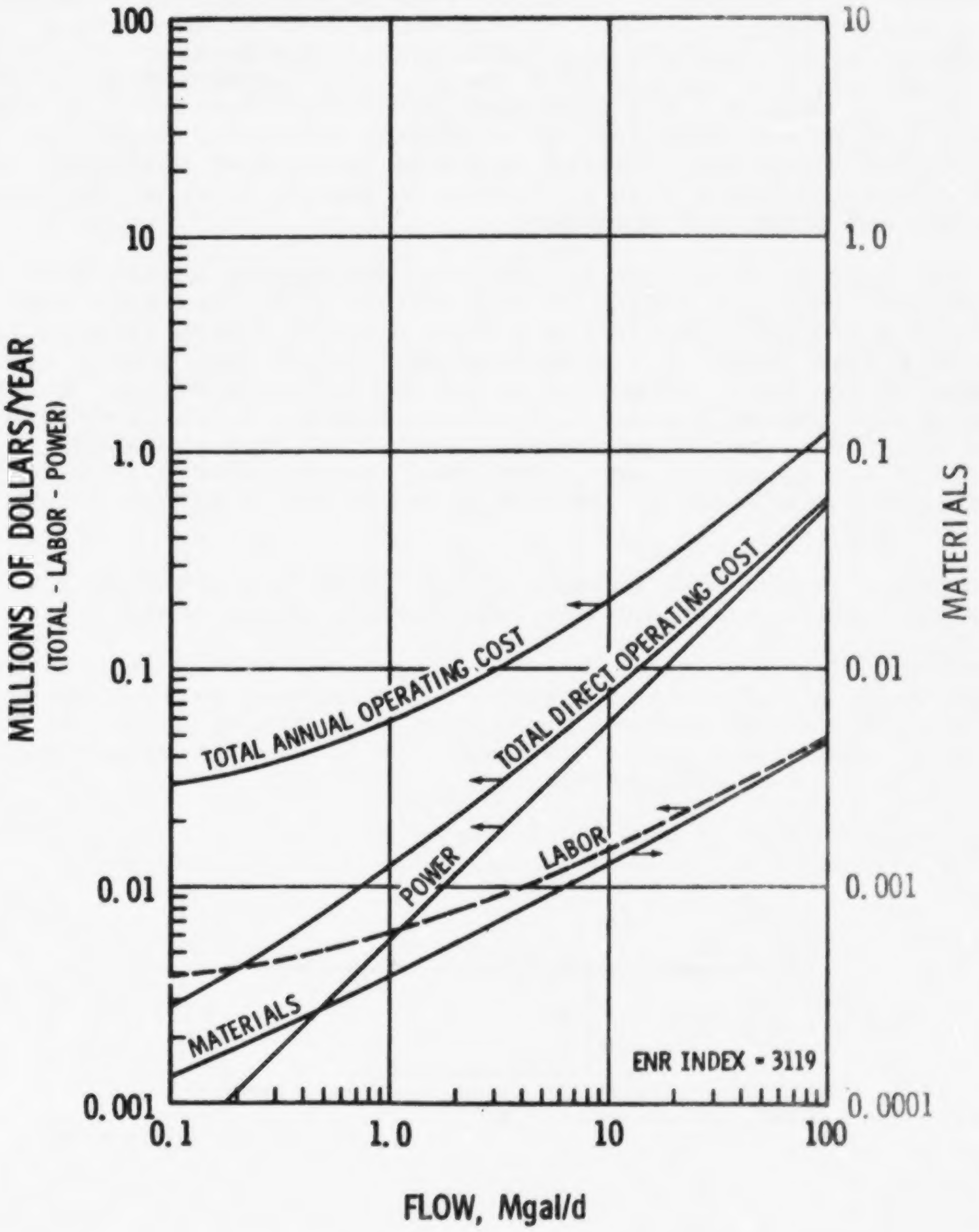
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



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IV.4.3.2 Anaerobic Lagoons [1]

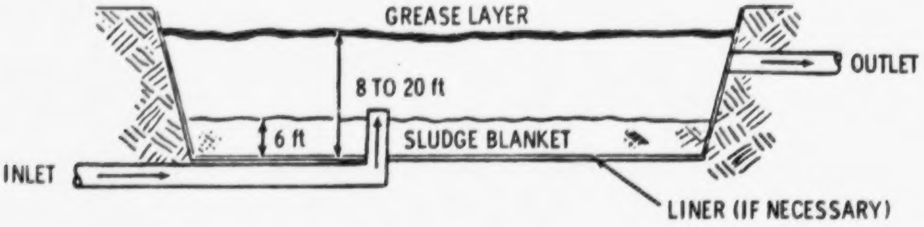
Description. Anaerobic lagoons are relatively deep (up to 20 ft) ponds with steep sidewalls in which anaerobic conditions are maintained by keeping loading so high that complete deoxygenation is prevalent. Although some oxygenation is possible in a shallow surface zone, once greases form an impervious surface layer, complete anaerobic conditions develop. Treatment or stabilization results from thermophilic anaerobic digestion of organic wastes. The treatment process is analogous to that occurring in the single-stage untreated anaerobic digestion of sludge in which acid-forming bacteria break down organics. The resultant acids are then converted to carbon dioxide, methane, cells, and other end products.

In the typical anaerobic lagoon, raw wastewater enters near the bottom of the pond (often at the center) and mixes with the active microbial mass in the sludge blanket, which is usually about 6 feet deep. The discharge is located near one of the sides of the pond, submerged below the liquid surface. Excess undigested grease floats to the top, forming a heat-retaining and relatively air-tight cover. Wasteflow flow equalization and heating are generally not practiced. Excess sludge is washed out with the effluent. Recirculation of waste sludge is not required.

Anaerobic lagoons are capable of providing treatment of high strength wastewaters and are resistant to shock loads.

Anaerobic lagoons are customarily contained within earthen dikes. Depending on soil characteristics, lining with various impervious materials such as rubber, plastic or clay may be necessary. Pond geometry may vary, but surface-area-to-volume ratios are minimized to enhance heat retention.

Typical equipment includes lining systems and hydraulic controls. A flow diagram for anaerobic lagooning is shown below.



Typical design criteria. Typical design criteria for anaerobic lagoons are shown below.

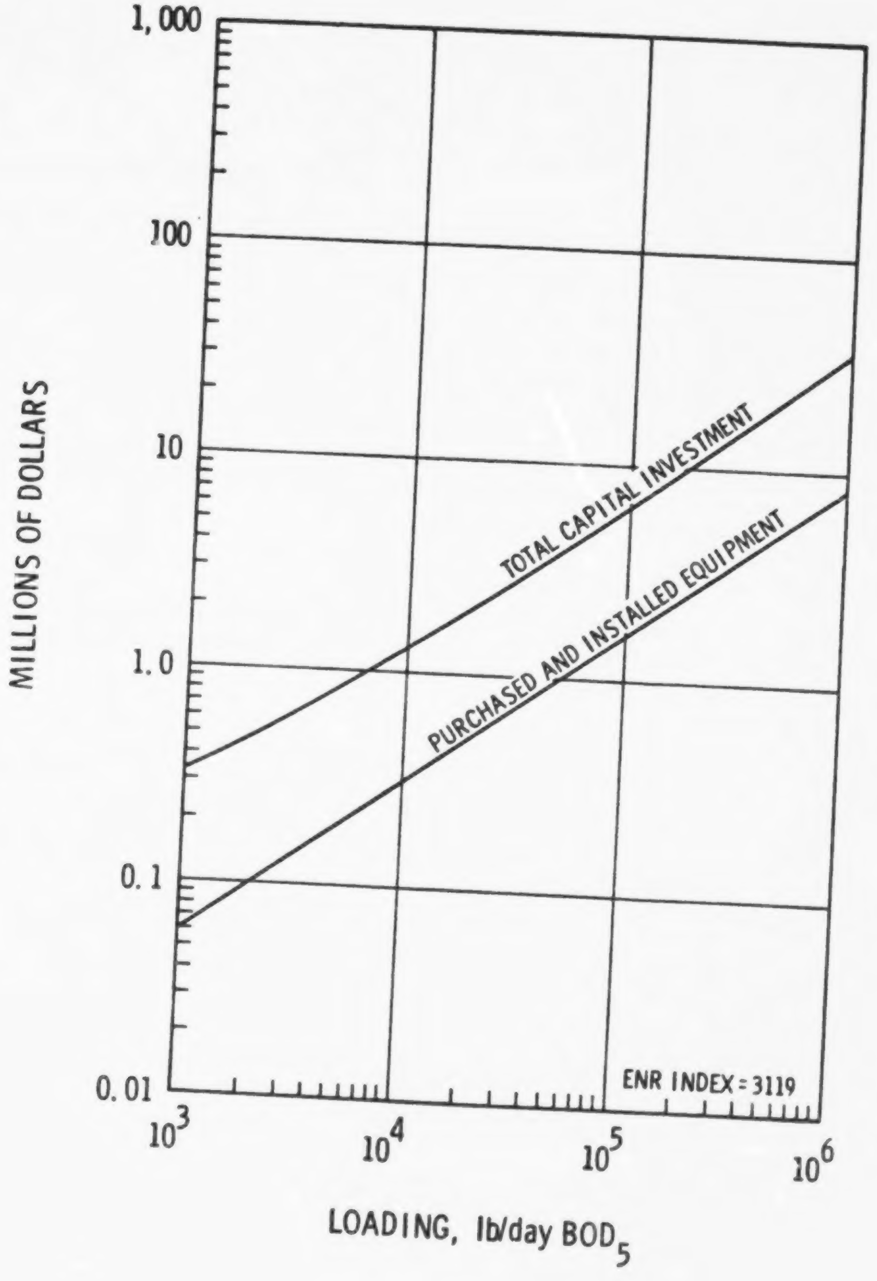
Design criteria	Value
Operation	Parallel
Detention time	1-50 d
Depth	8-20 ft
pH	6.8-7.2
Water temperature range	43-120°F
Optimum water temperature	86°F
Organic loading	220-2,200 lb BOD_5/acre/d

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes excavating, grading, other earth work, and service roads. These costs do not include land and pumping. Anaerobic lagoons are operated by gravity flow and, therefore, have no energy requirements other than any pumping that may be necessary to lift the influent wastewater into the lagoons. Liner cost is not included in the estimation. An average detention time of 35 days was assumed for cost estimation.

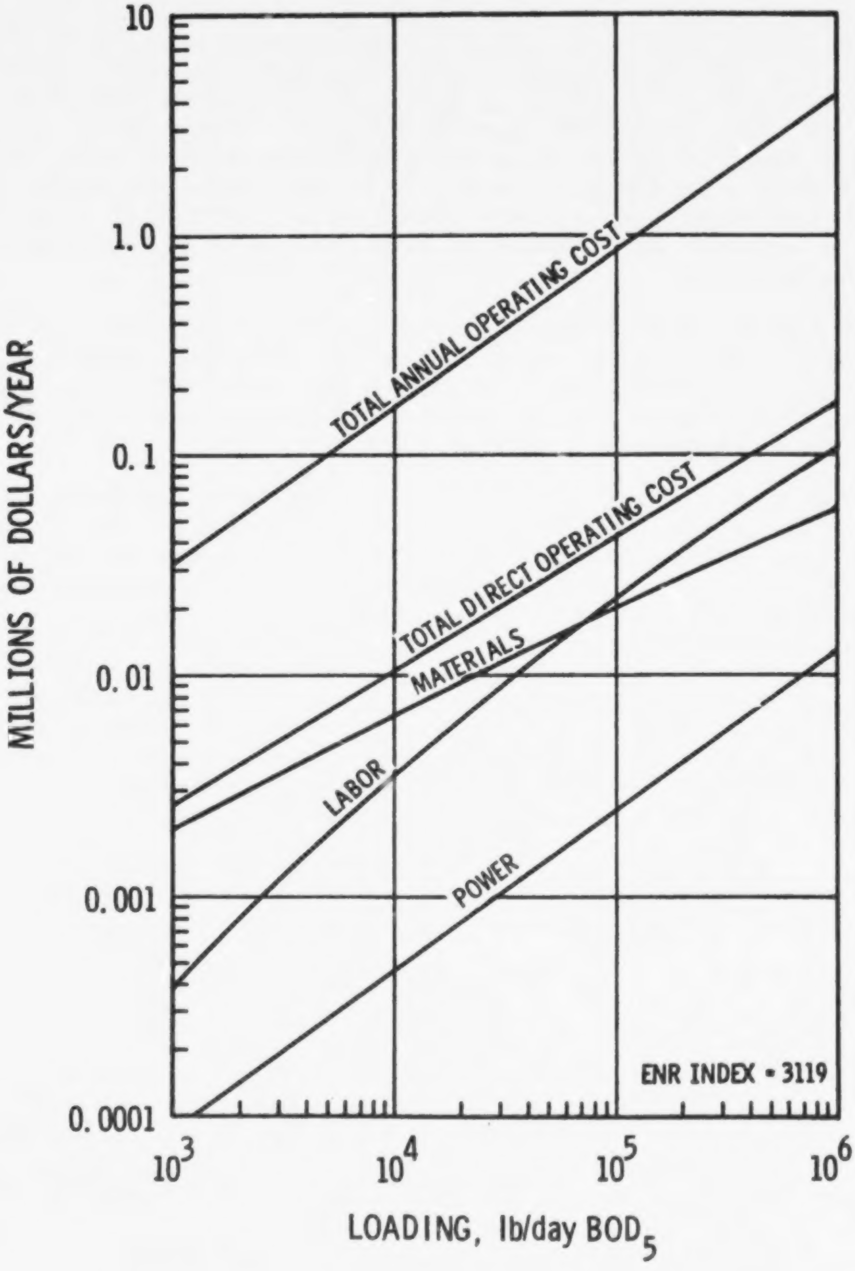
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.4.3.3 Facultative Lagoons [1]

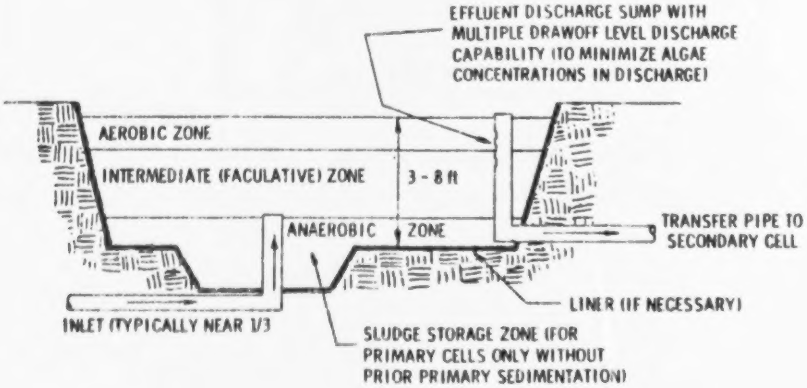
Description. Facultative lagoons are intermediate depth (3 to 8 feet) ponds in which the wastewater is stratified into three zones. These zones consist of an anaerobic bottom layer, an aerobic surface layer, and an intermediate zone. Stratification is a result of solids settling and temperature-water density variations. Oxygen in the surface stabilization zone is provided by reaeration and photosynthesis. This is in contrast to aerated lagoons in which mechanical aeration is used to create aerobic surface conditions. In general, the aerobic surface layer serves to reduce odors while providing treatment of soluble organic byproducts of the anaerobic process operating at the bottom.

Sludge at the bottom of facultative lagoons will undergo anaerobic digestion producing carbon dioxide, methane, and cells. The photosynthetic activity at the lagoon surface produces oxygen diurnally, increasing the dissolved oxygen during daylight hours, while surface oxygen is depleted at night.

Facultative lagoons are often, and for optimum performance, should be, operated in series. When three or more cells are linked, the effluent from either the second or third cell may be recirculated to the first. Recirculation rates of 0.5 to 2.0 times the plant flow have been used to improve overall performance.

Facultative lagoons are customarily contained within earthen dikes. Depending on soil characteristics, lining with various impervious materials such as rubber, plastic or clay may be necessary. Use of supplemental top-layer aeration can improve overall treatment capacity, particularly in northern climates where icing over of facultative lagoons is common in the winter.

Typical equipment includes lining systems and hydraulic controls. A flow diagram for a facultative lagoon is shown below.


EFFLUENT DISCHARGE SUMP WITH MULTIPLE DRAWOFF LEVEL DISCHARGE CAPABILITY (TO MINIMIZE ALGAE CONCENTRATIONS IN DISCHARGE)
AEROBIC ZONE
INTERMEDIATE (FACULATIVE) ZONE
3 - 8 ft
ANAEROBIC ZONE
TRANSFER PIPE TO SECONDARY CELL
LINER (IF NECESSARY)
INLET (TYPICALLY NEAR 1/3
SLUDGE STORAGE ZONE (FOR PRIMARY CELLS ONLY WITHOUT PRIOR PRIMARY SEDIMENTATION)

Typical design criteria. Typical design criteria for facultative lagoons are shown below.

Design criteria	Value
Operation	At least three cells in series are required; parallel trains of cells may be used for larger systems
Detention time	20-180 d
Depth	3-8 ft, although a portion of the anaerobic zone of the first cell may be up to 12 ft deep to accommodate large initial solids deposition
pH	6.5-9.0
Water temperature range	35-90°F for municipal applications
Optimum water temperature	68°F
Organic loading	10-100 lb BOD_5/acre/d

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes excavating, grading, and other earth work required for normal subgrade preparation, and service roads. These costs do not include land, pumping, and liner. Facultative lagoons are operated by gravity, and, therefore, have no energy requirements other than any pumping that may be necessary to lift the influent wastewater into the lagoons. The following operating and wastewater characteristics were assumed for cost estimation:

Operating/wastewater characteristic	Assumed value
Lagoon loading	40 lb BOD_5/acre/d in warm climate; 20 lb BOD_5/acre/d in cool climate
Water depth	4 ft
BOD_5	210 mg/L influent; 30 mg/L effluent
COD	400 mg/L influent; 100 mg/L effluent
TSS	230 mg/L influent; 60 mg/L effluent
Total-P	11 mg/L influent; 8 mg/L effluent
NH_3-N	20 mg/L influent; 15 mg/L effluent in cool climate and 15 mg/L in warm climate

To adjust costs for loadings other than those above, calculate the effective flow (Q_E) using the following equations:

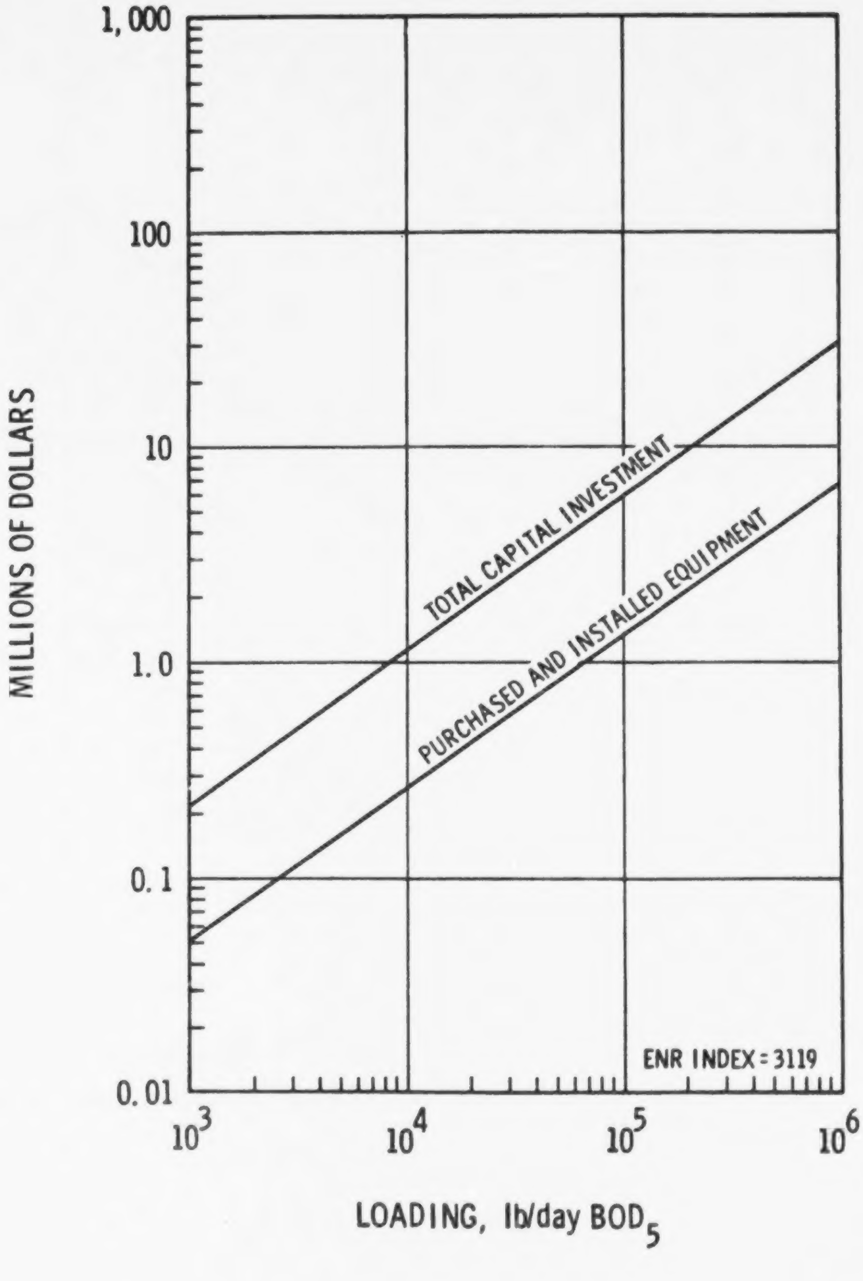
$$\text{For warm climate } Q_E = Q_{Design} \times \frac{40\ \text{lb BOD}_5\text{/acre/day}}{\text{new design loading}}$$

$$\text{For cold climate } Q_E = Q_{Design} \times \frac{20\ \text{lb BOD}_5\text{/acre/day}}{\text{new design loading}}$$

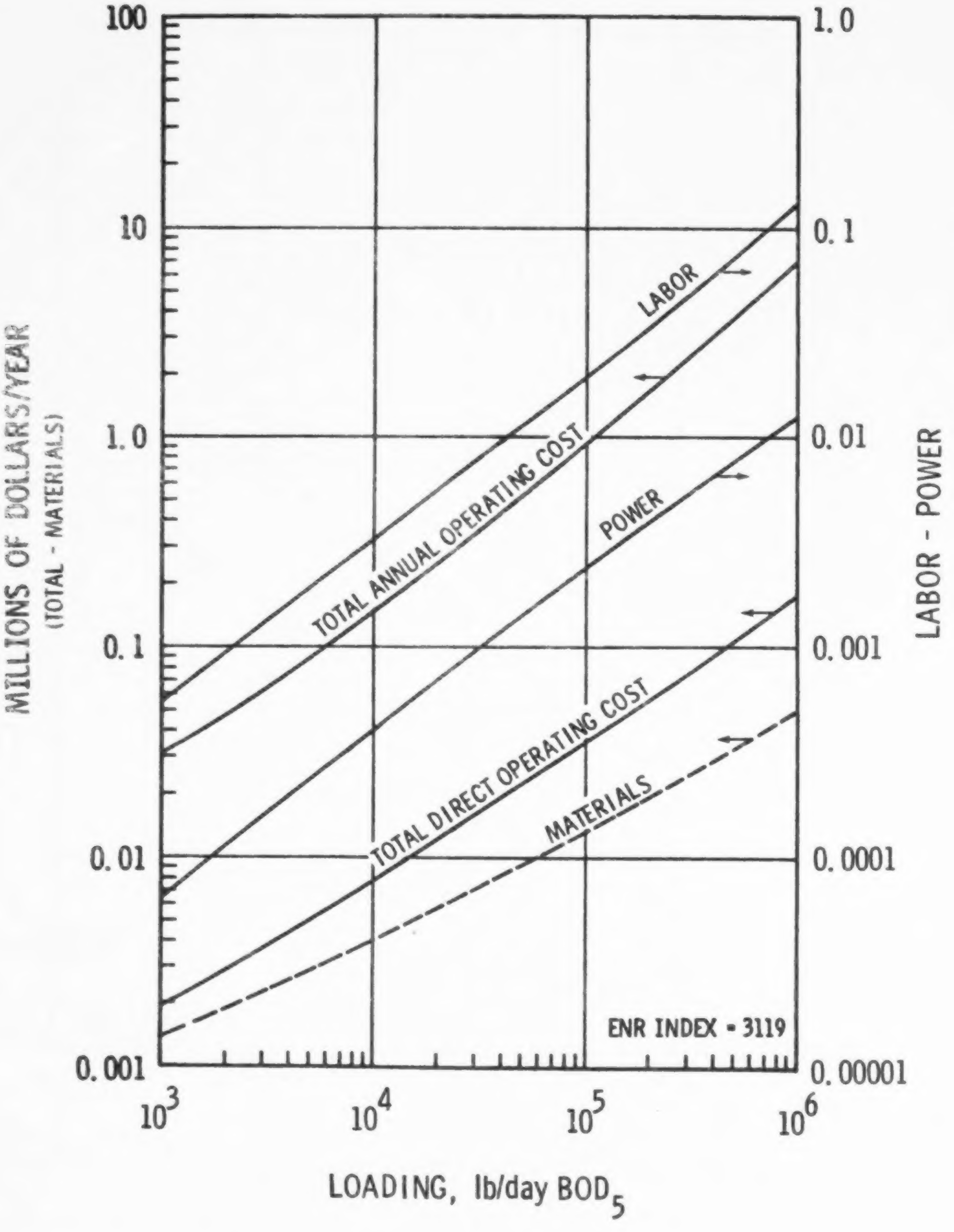
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



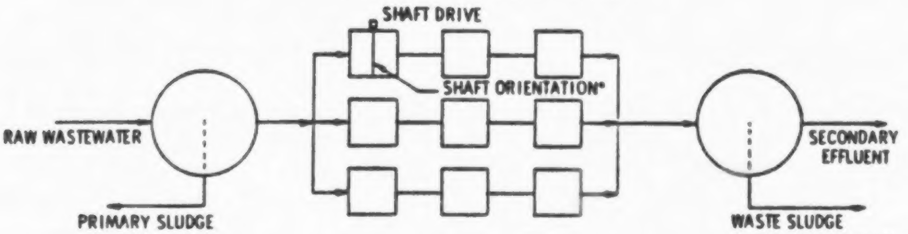
IV.4.4 ROTATING BIOLOGICAL CONTACTORS [1]

IV.4.4.1 Description

Rotating biological contactors (RBC) are used to remove dissolved and collodial biodegradable organics. The process utilizes a fixed-film biological reactor consisting of plastic media mounted on a horizontal shaft and placed in a tank. Common media forms are a disc-type made of styrofoam and a denser lattice-type made of polyethylene. While wastewater flows through the tank, the media are slowly rotated, about 40% immersed, for contact with the wastewater to remove organic matter by the biological film that develops on the media. Rotation results in exposure of the film to the atmosphere as a means of aeration. Excess biomass on the media is stripped off by rotational shear forces, and the stripped solids are maintained in suspension by the mixing action of the rotating media. Multiple staging of RBC's increases treatment efficiency and could aid in achieving nitrification year round. A complete system could consist of two or more parallel trains, each consisting of multiple stages in series. Common modifications of RBC's include the following: multiple staging; use of dense media for latter stages in train; use of molded covers or housing of units; various methods of pretreatment and after treatment of wastewater; use in combination with trickling filter or activated sludge processes; use of air driven system in lieu of mechanically driven system; addition of air to the tanks; addition of chemicals for pH control; and sludge recycle to enhance nitrification.

Equipment normally associated with rotating biological contactors includes rotating disc systems. A flow diagram of a typical staged rotating biological contactor configuration is shown below.

TYPICAL STAGED RBC CONFIGURATION



(SEE SECTION IV.6)

(SEE SECTION IV.6)

*ALTERNATE SHAFT ORIENTATION IS PARALLEL TO DIRECTION OF FLOW WITH A COMMON DRIVE FOR ALL THE STAGES IN A SINGLE TRAIN.

IV.4.4.2 Typical Design Criteria

Typical design criteria of rotating biological contactors are tabulated on the following page.

Design criteria	Value
Organic loading T >55°F	Without nitrification: 4-5 - 60 lb BOD_5/1,000 ft^2 of media/day With nitrification: 1.5-3.0 lb BOD_5/1,000 ft^3 of media
Hydraulic loading	Without nitrification: 1.5 - 3 gpd/ft^2 of media surface area With nitrification: 0.9 - 2 gpd/ft^3 of media surface area
Number of stages/ train	At least 4
Number of parallel trains	At least 2
Rotational velocity	60 ft/min (peripheral, mechanical drive)
Media surface area	Disc type: 20 - 25 ft^2/ft^3 Lattice type: 30 - 35 ft^2/ft^3
Media submerged	40%
Tank volume	0.12 gal/ft^2 of disc area
Detention time	Without nitrification: 55 - 120 min (based on 0.12 gal/ft^2) With nitrification: 80 - 190 min (based on 0.12 gal/ft^2)
Secondary clarifier overflow rate	500 - 700 gpd/ft^2
Power	7 - 20 hp/25 ft shaft

IV.4.4.3 Costs

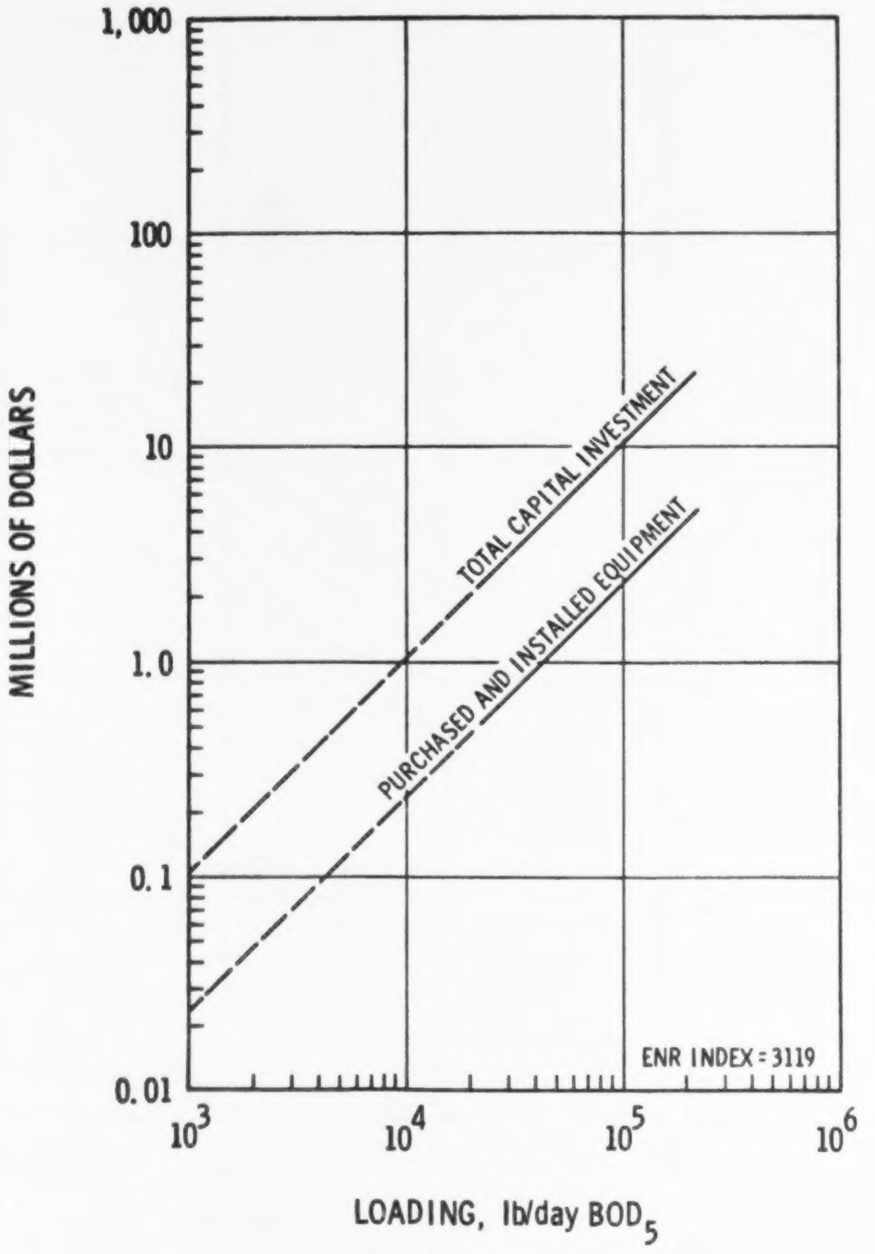
Purchased equipment and installation cost for estimation of the total capital investment includes RBC shafts (standard media, 100,000 ft^2/shaft), motor drives (5 hp/shaft), molded fiberglass covers, and reinforced concrete basins. Final clarifiers are not included in the purchased and installed equipment cost. Cost is based on organic loading rate of 2,800 mg/L.

$$Q_E = Q_{Design} \times \frac{1.0\ \text{gpd/ft}^2}{\text{Actual hydraulic loading (gpd/ft}^2\text{)}}$$

IV.4.4.4 References

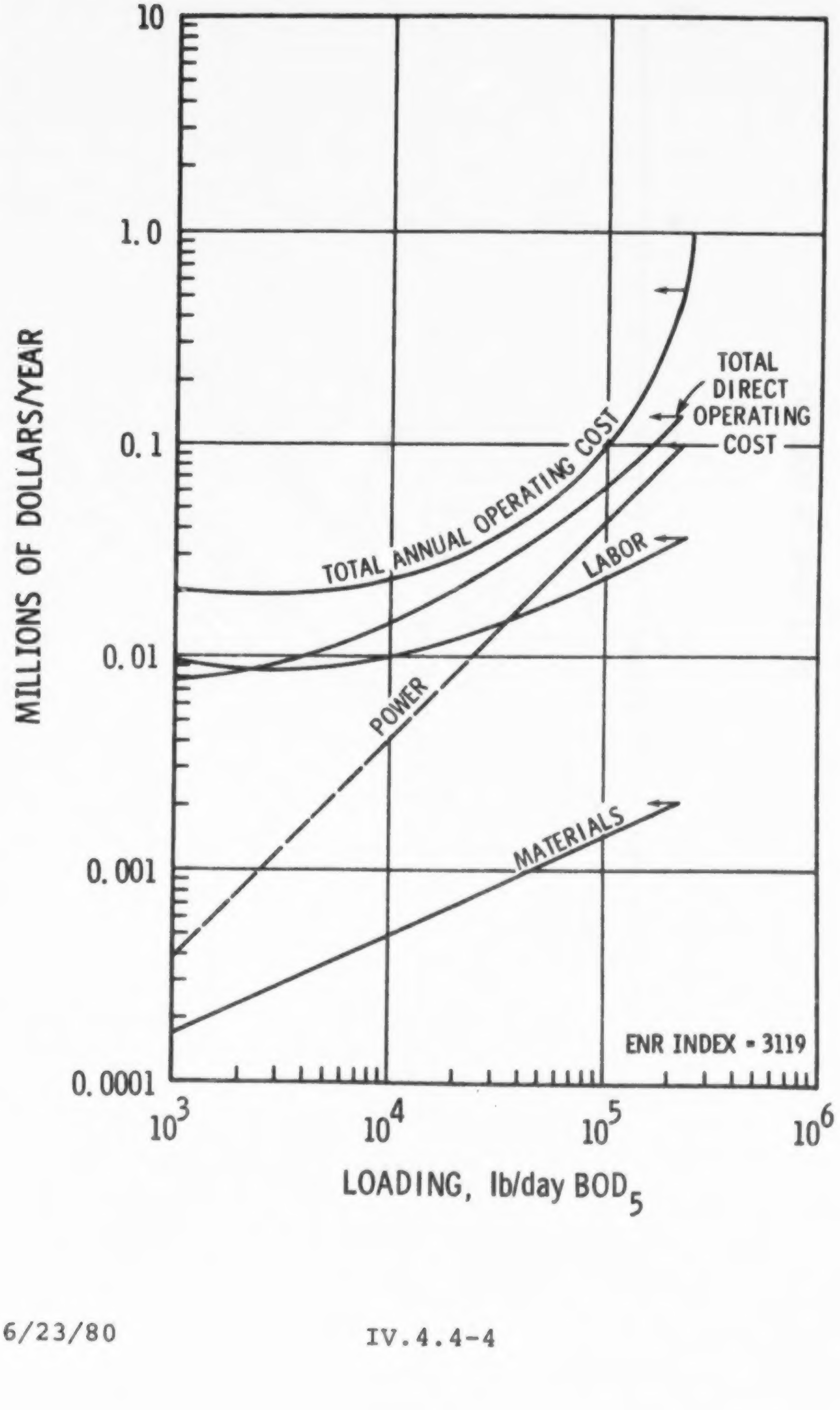
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



Date: 6/23/80

IV.4.5 STEAM STRIPPING [1]

IV.4.5.1 Description

Steam stripping is essentially a fractional distillation of volatile compounds from a wastewater stream. The volatile component may be a gas or volatile organic compound with solubility in the wastewater stream. In most instances, the volatile component, such as methanol or ammonia, is quite water soluble.

Steam stripping is usually conducted as a continuous operation in a packed tower or conventional fractionating distillation column (bubble cap or sieve tray) with more than one stage of vapor/liquid contact. The preheated wastewater from the heat exchanger enters near the top of the distillation column and then flows by gravity countercurrent to the steam and organic vapors (or gas) rising up from the bottom of the column. As the wastewater passes down through the column, it contacts the vapors rising from the bottom of the column that contain progressively less volatile organic compound or gas. When the wastewater reaches the bottom of the column, it is heated by the incoming steam to reduce the concentration of volatile component(s) to their final concentration. Much of the heat in the wastewater discharged from the bottom of the column is recovered in preheating the feed to the column.

Reflux (i.e., condensing a portion of the vapors from the top of the column and returning it to the column) may or may not be practiced depending on the composition of the vapor stream that is desired. Although many of the steam strippers in industrial use introduce the wastewater at the top of the stripper, there are advantages to introducing the feed to a tray below the top tray when reflux is used.

Introducing the feed at a lower tray (while still using the same number of trays in the stripper) will have the effect of either reducing steam requirements (due to the need for less reflux) or yielding a vapor stream richer in volatile component). The combination of using reflux and introducing the feed at a lower tray will increase the concentration of the volatile organic component beyond that obtainable by reflux alone.

Equipment for steam stripping is nearly the same as that required for conventional fractional distillation (i.e., packed column or tray tower, reboiler, reflux condenser and feed tanks, and pumps). However, a heat exchanger is used for heating feed entering column and cooling stripped wastewater leaving column; the reboiler is often an integral part of the tower body rather than a separate vessel; materials of construction depend on operating pH and presence (or absence) of corrosive ions (i.e., sulfides, chlorides); in a single-column sour-water steam stripper, the high pH (from the presence of ammonia) allows use of mild steel; if sour water is stripped in two columns (H_2S removed in one and NH_3 removed in other), alloy steel or alloy-clad steel should be used in unit in which H_2S is removed.

IV.4.5.2 Typical Design Criteria

Design criteria	Value
Wastewater feed	250 ml/min; 3.8 L/m design
Steam feed rate	59.7 mL/min
Volumetric flowrate:	
overhead	4.3 mL/min
bottoms	305 mL/min
Temperature feed	
overhead	104°C
bottoms	104°C
Column height	3.67 m
Column diameter	5.08 cm

IV.4.5.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes feed pumps (130-ft head and 60-ft head), feed heat exchanger (800 ft^2), distillation column (24 trays, 6-ft diameter x 60 ft), condenser (400 ft^2), and condensate tank (1,000 gal). The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Wastestream	0.288 Mgpd sour water containing 5% $(NH_4)_2S$ (by wt)
Operation	350 d/yr, 24 hr/d

Cost item	Cost per unit quantity	Annual cost, $	
Direct operating cost			
Labor			
Operating	$16/man-hr	33,600	
Maintenance		28,000	
			51,600
Materials			29,500
Steam	$5/1,000 lb		1,008,000
Cooling water	$0.043/1,000 gal		26,000
Power	$0.035/kWh		6,000
Total			1,121,000
Total indirect operating cost			198,000
Total annual operating cost			1,320,000

Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying flowrates:

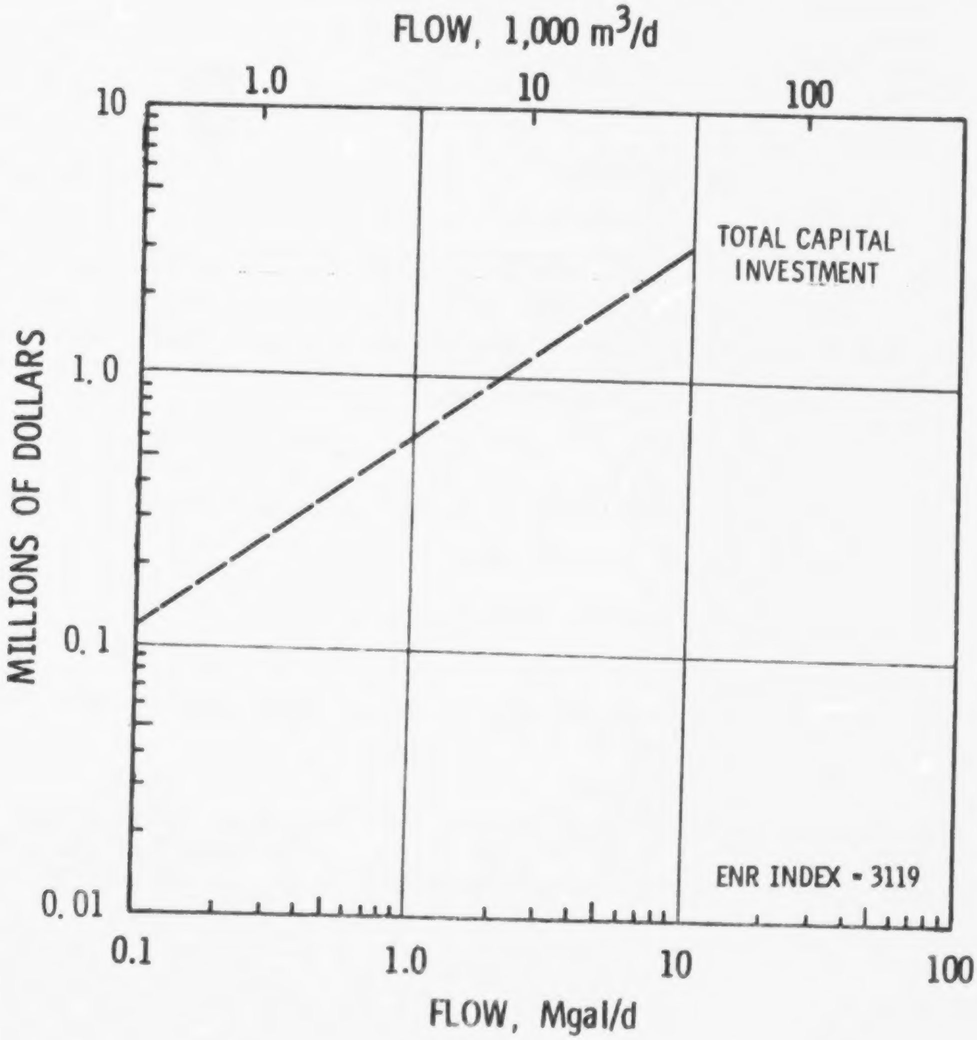
Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Materials, power, and chemicals	1.0

The capital cost for 106 MM gal/yr (0.286 Mgal/d) is $850,000.

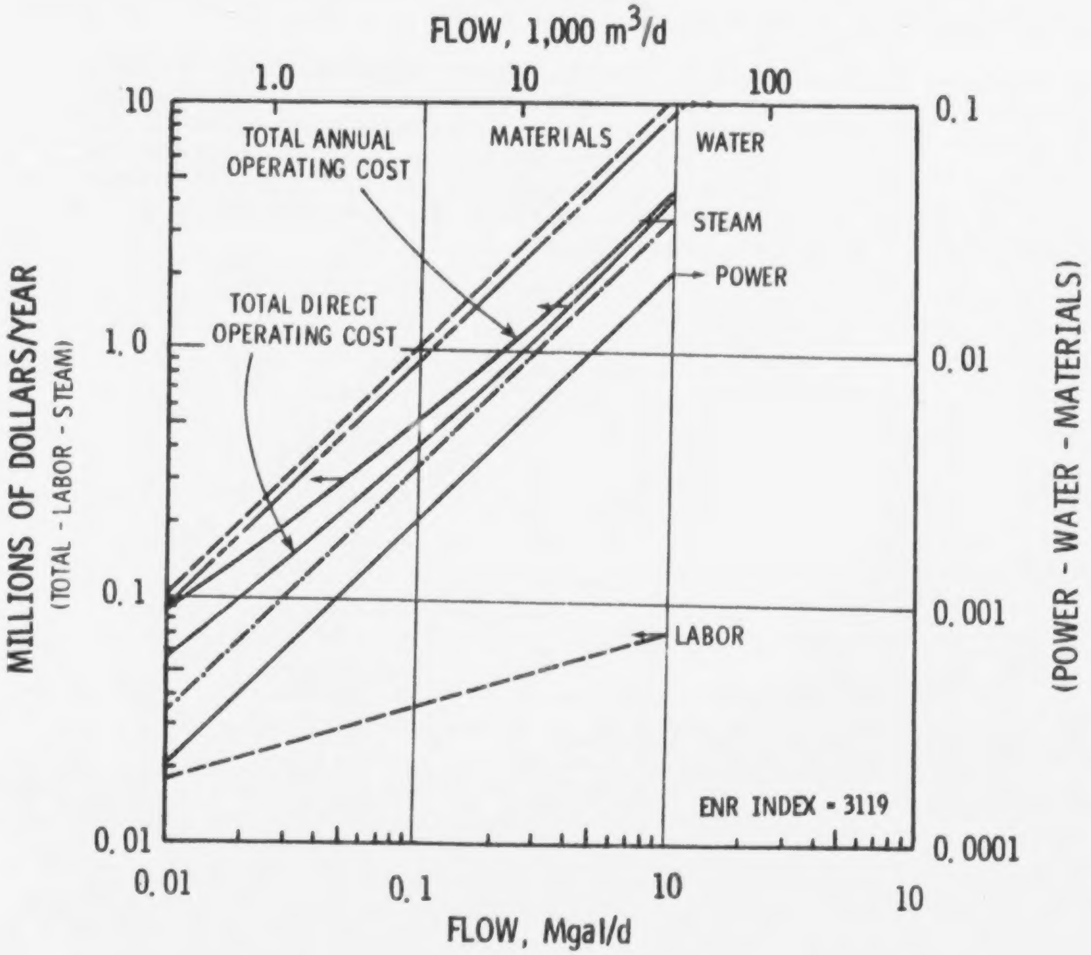
IV.4.5.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST

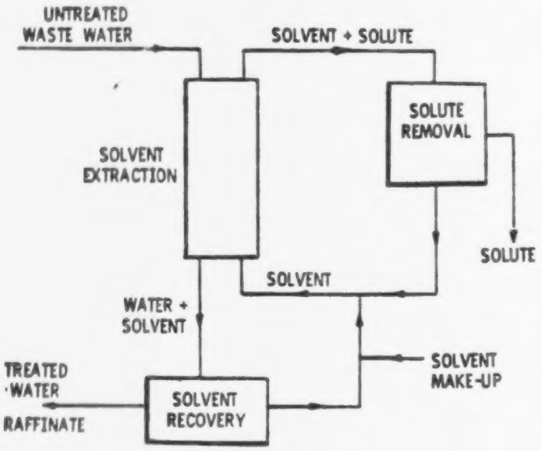


IV.4.6 SOLVENT EXTRACTION [1]

IV.4.6.1 Description

Liquid-liquid solvent extraction, hereinafter referred to as solvent extraction, is the separation of the constituents of a liquid solution by contact with another immiscible liquid. If the substances comprising the original solution distribute themselves differently between the two liquid phases, a certain degree of separation will result, and this may be enhanced by use of multiple contacts.

The solvent extraction process is shown schematically below. The diagram shows a single solvent extraction unit operating on an aqueous stream, in practice this unit might consist of (1) a single-stage mixing and settling unit, (2) several mixers and settlers (single-stage unit) in series, or (3) a multi-stage unit operating by countercurrent flows in one device (e.g., a column or differential centrifuge).



As indicated in the diagram, reuse of the extracting solvent (following solute removal) and recovery of that portion of the extracting solvent that dissolves in the extracted phase are usually necessary aspects of the solvent extraction process. Solvent reuse is necessary for economic reasons; the cost of the solvent is generally too high to consider disposal after use. Only in a very few cases may solvent reuse be eliminated; these cases arise when an industrial chemical feed stream can be used as the solvent and then sent on for normal processing, or where water is the solvent. Solvent recovery from extracted water may be eliminated in cases when the concentration in the water to be discharged is not harmful, and when the solvent loss does not incur a high cost.

The end result of solvent extraction is to separate the original solution into two streams: a treated stream (the raffinate), and a recovered solute stream (which may contain small amounts of water and solvent). Solvent extraction may thus be considered a recovery process since the solute chemicals are generally recovered for reuse, resale, or further treatment and disposal.

A process for solvent extracting a solution will typically include three basic steps: the actual extraction, solute removal from the extracting solvent, and solvent recovery from the raffinate (treated stream). The process may be operated continuously.

The first step, extraction, brings the two liquid phases (feed and solvent) into intimate contact to allow solute transfer either by forced mixing or by countercurrent flow caused by density differences. The extractor will also have provisions to allow separation of the two phases after mixing. One output stream from the extractor is the solute-laden solvent; some water may also be present. Solute removal may be via a second solvent extraction step, distillation, or some other process. For example, a second extraction, with caustic, is sometimes used to extract phenol from light oil, which is used as the primary solvent in diphenolizing coke plant wastewaters. Distillation will usually be more common, except when problems with azeotropes are present. In certain cases, it may be possible to use the solute-laden solvent as a feed stream in some industrial process, thus eliminating solute recovery. This is apparently the case at some refineries where crude or light oil can be used as a solvent (for phenol removal from water) and later processed with the solute in it. Other similar applications probably exist and are particularly attractive because they eliminate one costly step. Solvent recovery from the treated stream may be required if solvent losses would otherwise add significantly to the cost of the process, or cause a problem with the discharge of the raffinate. Solvent recovery may be accomplished by stripping, distillation, adsorption, or other suitable processes.

There are two major categories of equipment for liquid extraction: single-stage and multi-stage equipment.

In single-stage equipment, the fluids are mixed, extraction occurs, and the insoluble liquids are settled and separated. A cascade of such stages may then be arranged. A single-stage must provide facilities for mixing the insoluble liquids and for settling and decanting the emulsion or dispersion which results. In batch operation, mixing together with settling and decanting may take place in the same or in separate vessels. In continuous operation, different vessels are required.

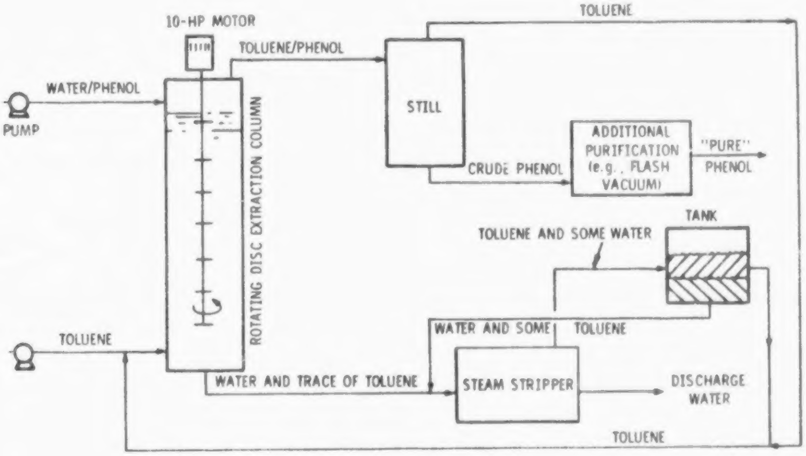
In multi-stage equipment, the equivalent of many stages may be incorporated into a single device or apparatus. Countercurrent flow is produced by virtue of the difference in densities of the liquids, and with few exceptions the equipment takes the form of a vertical tower which may or may not contain internal devices to influence the flow pattern. Other forms include centrifuges, rotating discs, and rotating buckets. Depending upon the nature of the internal structure, the equipment may be of the stagewise or continuous-contact type.

IV.4.6.2 Typical Design Criteria

Design criteria	Value
Column diameter	0.0762 m
Column length (glass pipe)	1.22 m
Solvent flow rate	0.457 m/hr
Wastewater flowrate	3.60 m/hr

IV.4.6.3 Costs

It is quite difficult to generalize about the costs of solvent extraction because of the wide variety of systems, feed streams, and equipment that may be involved. In order to present a generalized cost estimate, however, data are generated for a specific application; i.e., the removal of phenol from water with toluene by solvent extraction. A flow diagram of this system and its operating characteristics are shown below.


10-HP MOTOR
TOLUENE/PHENOL
TOLUENE
WATER/PHENOL
PUMP
STILL
ROTATING DISC EXTRACTION COLUMN
ADDITIONAL PURIFICATION (e.g., FLASH VACUUM)
"PURE" PHENOL
CRUDE PHENOL
TANK
TOLUENE AND SOME WATER
WATER AND SOME
TOLUENE
TOLUENE
WATER AND TRACE OF TOLUENE
STEAM STRIPPER
DISCHARGE WATER
TOLUENE

Operating characteristics	Value
Water/phenol/feed	45,000 lb/hr containing 1.5% phenol (by wt); temperature is 110°F.
Toluene feed	45,000 lb/hr (containing ∿20 ppm phenol from recycle)
Discharge water	Contains 75 ppm phenol
Extraction column	Rotating disc type; 6-ft diameter, 60 ft high; made from carbon steel; contains 50 compartments and equivalent of about 5 theoretical stages. (Equilibrium distribution coefficient of phenol between toluene and water is about 2.)
Loss of toluene/cycle	Approximately 0.1%/cycle
Electrical requirements (column only)	One 10 hp electric motor
Operation	330 d/yr; 24 hr/d

Fixed capital costs based on an ENR index of 3119 are estimated to be $1,100,000. Similarly, total capital cost is estimated to be $1,150,000. Estimation of annual operating cost is presented in the following table.

Cost item	Annual quantity	Cost per unit quantity	Annual cost,[a] $	
Direct operating cost				
Labor				
Operating	12,000 man-hr	$16/hr	192,000	
Maintenance			16,000	
				208,000
Chemicals - Toluene	15,000 gal	$1.15/gal		17,300
Materials				16,700
Steam	33 x 10^6 lb	$5/1,000 lb		165,000
Power	150,000 kWh	$0.035/kWh		5,300
Total				412,000
Total indirect operating cost				350,000
Total annual operating cost[a]				762,000

[a] Excludes annual credit for phenol recovery.

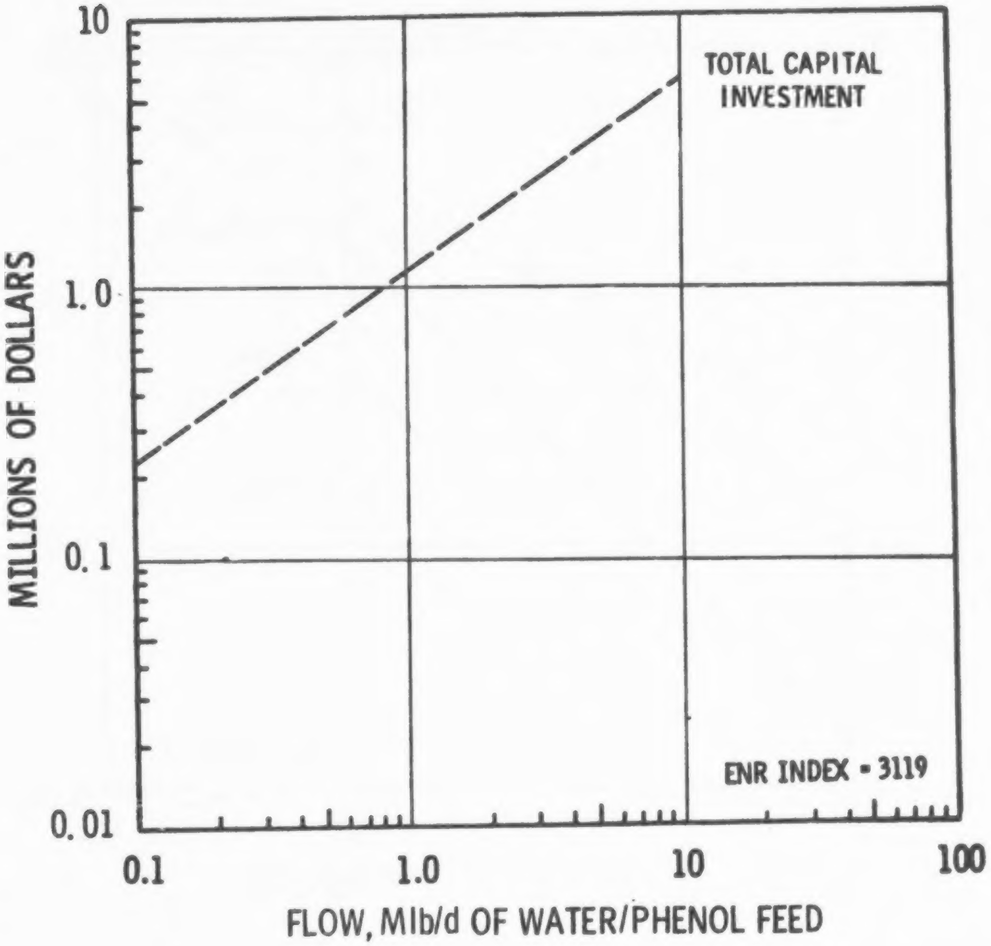
Cost curves were developed for total capital investment and annual operating cost using the cost data shown on the previous page and the following exponential scaling factors, which were used to determine cost at varying capacities:

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power, materials, steam and chemicals	1.0

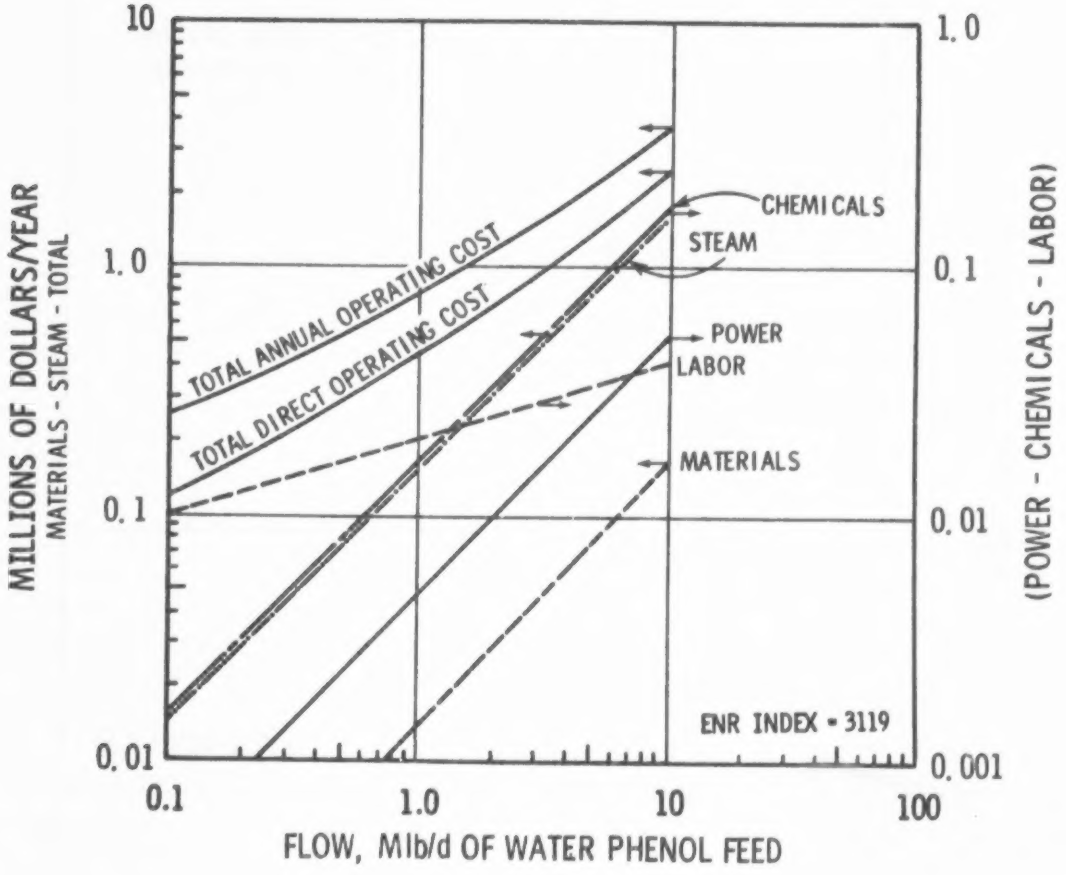
IV.4.6.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C. November 1976. pp. 32-1 through 30-26.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



Date: 6/23/80

IV.5 TERTIARY WASTEWATER TREATMENT

IV.5.1 GRANULAR ACTIVATED CARBON ADSORPTION [1]

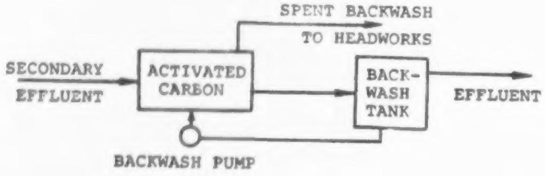
IV.5.1.1 Description

Activated carbon adsorption is used for the removal of dissolved organics and control of such wastewater parameters as COD, TOC, BOD_5, TOD, and specific soluble organic materials. In most cases, activated carbon is used as an individual stream pretreatment process; however, in other cases activated carbon treatment is used as a final treatment process following biological treatment.

Granular carbon systems generally consist of vessels in which the carbon is placed, forming a "filter" bed. These systems can also include carbon storage vessels and thermal regeneration facilities. Vessels are usually circular for pressure systems or rectantular for gravity flow systems. Once the carbon adsorptive capacity has been fully utilized, the carbon must be disposed of or regenerated. Usually multiple carbon vessels are used to allow continuous operation. Columns can be operated in a series or parallel mode. All vessels must be equipped with carbon removal and loading mechanisms to allow for the removal of spent carbon and the addition of new material. Flow can be either upward or downward through the carbon bed. Vessels are backwashed periodically. Surface wash and air scour systems can also be used as part of the backwash cycle.

Small systems usually dispose of spent carbon or regenerate it off site. Systems using above about 1,000 lb/d usually provide on-site regeneration of carbon for economic reasons.

Equipment normally associated with granular activated carbon adsorption systems includes activated carbon material and granular carbon systems. A flow diagram of the activated carbon adsorption process is shown below.



IV.5.1.2 Typical Design Criteria

Typical design criteria for an granular activated carbon adsorption process are shown below.

Design criterion	Value
Vessel size	2 - 12 ft diameter commonly used
Area loading	2 - 10 gpm/ft^2
Organic loading	0.1 - 0.3 lb BOD_5(or COD)/lb carbon
Backwash	12 - 20 gpm/ft^2
Air scour	3 - 5 ft^3/min/ft
Bed depth	5 - 30 ft
Contact time	10 - 50 min
Land area	Minimal

IV.5.1.3 Costs

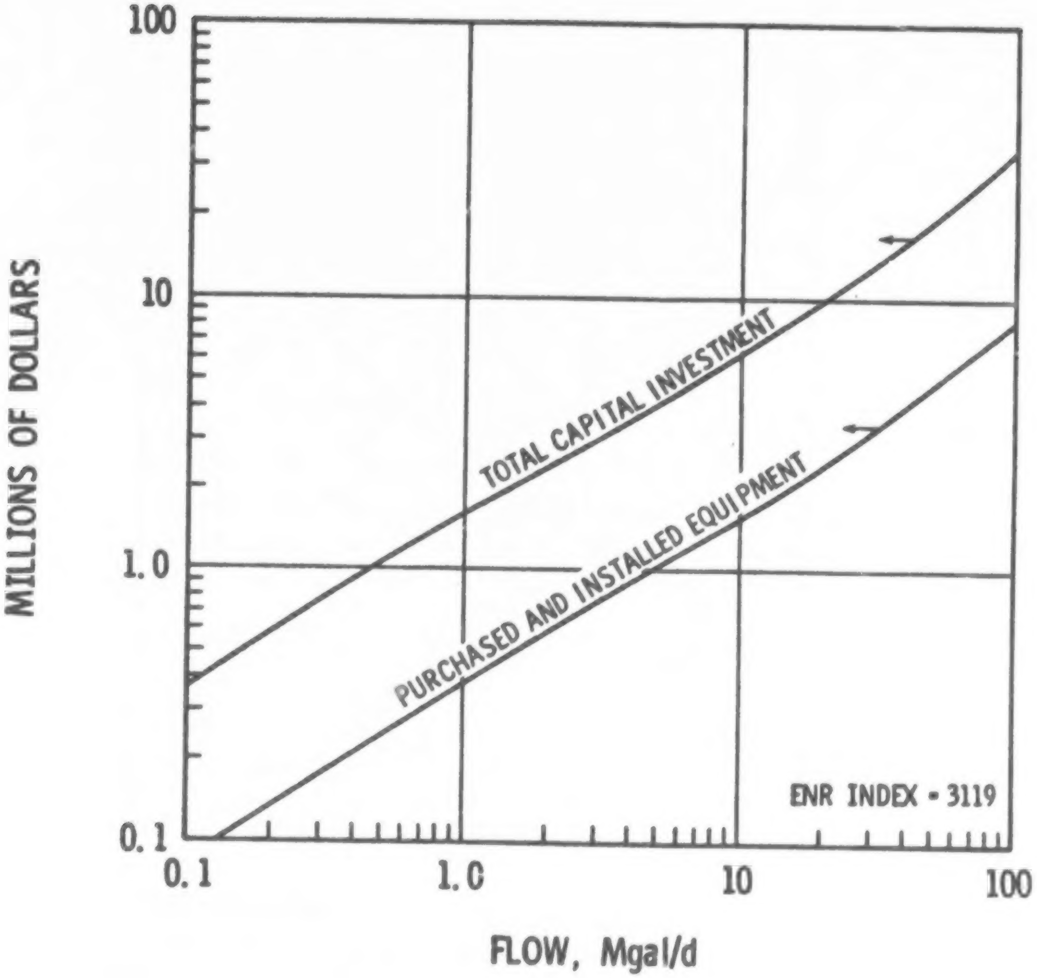
Purchased equipment and installation cost for estimation of the total capital investment includes vessels, media, pumps, carbon storage tanks, controls, and operations building. Disposal costs are not included. No regeneration is included; therefore, above 3 Mgal/d cost curves are extrapolated. The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Loading rate	30 lb/Mgal
Contact time	30 min

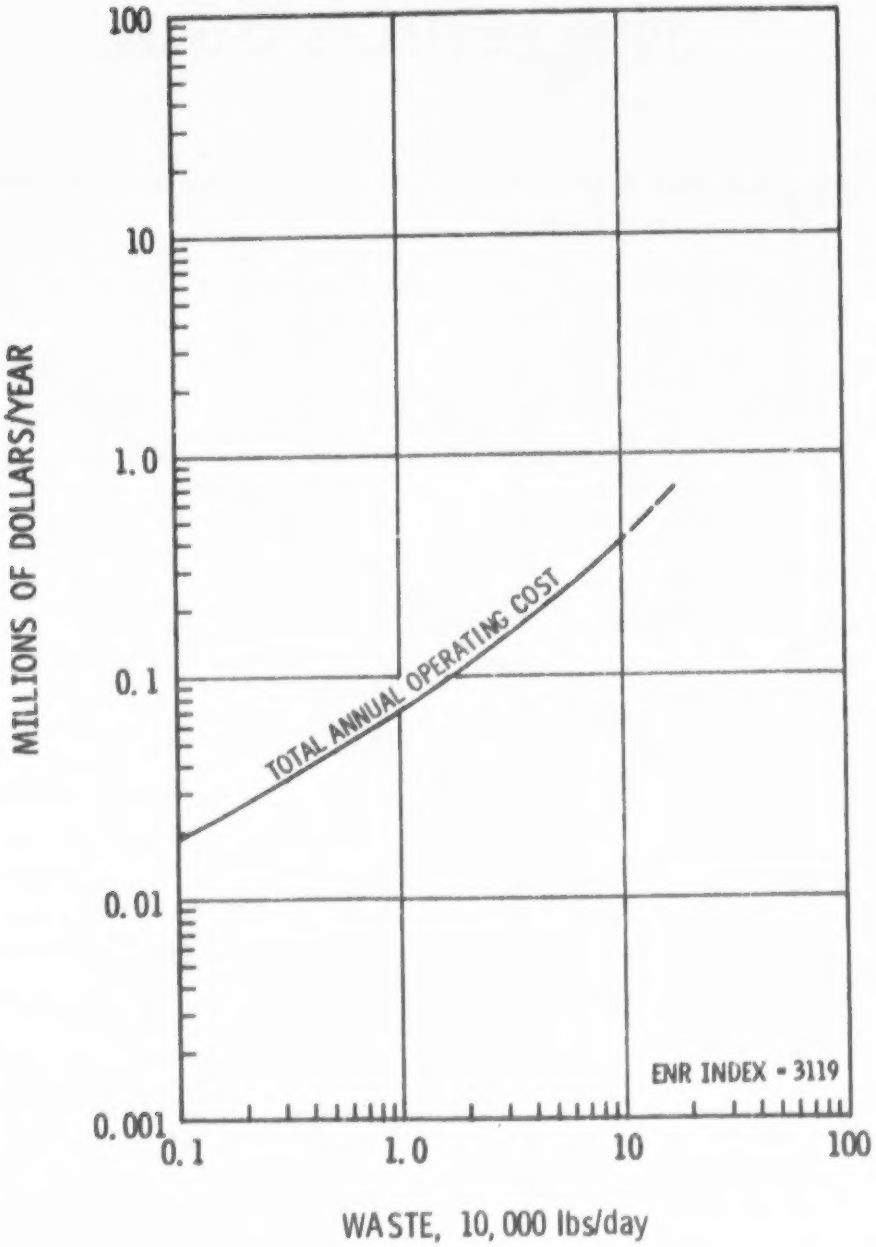
IV.5.1.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

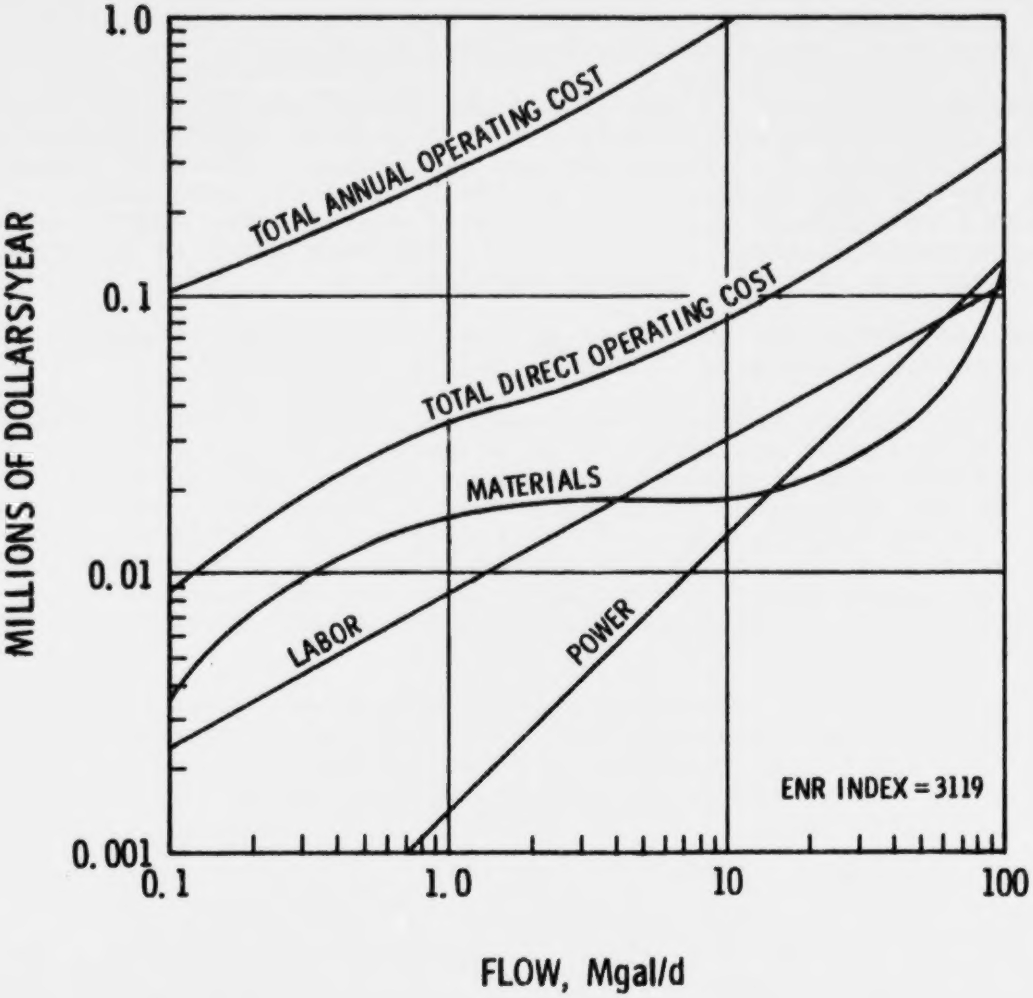
TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



ANNUAL OPERATING COST



IV.5.2 POWDERED ACTIVATED CARBON ADSORPTION [1]

IV.5.2.1 Description

Powdered activation carbon is used in wastewater facilities to absorb soluble organic materials and to aid in the clarification process.

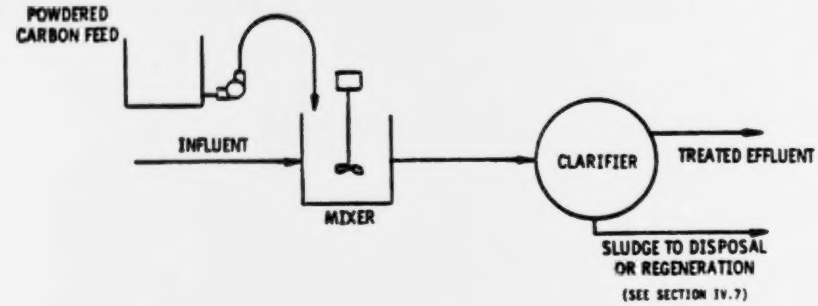
Powdered carbon is fed to a treatment system using chemical feed equipment similar to that used for other chemicals that are purchased in dry form. The spent carbon is removed with the sludge and then discarded or regenerated. Regeneration can be accomplished in a furnace or wet air oxidation system.

Powdered carbon can be fed to primary clarifiers directly, or to a separate sludge recirculation-type clarifier that enhances the contact between the carbon and the wastewater. Powdered carbon can also be fed to tertiary clarifiers to remove additional amounts of soluble organics. Powdered carbon, when added to a sludge recirculation-type clarifier, has been shown to be capable of achieving secondary removal efficiencies.

Powdered carbon can be fed in the dry state using volumetric or gravimetric feeders or it can be fed in slurry form.

A new technology has been developed over the past several years that consists of the addition of powdered activated carbon to the aeration basins of biological systems. This application is capable of the following: high BOD_5 and COD reduction, despite hydraulic and organic overloading; aiding solids settling in the clarifiers; a high degree of nitrification due to extended sludge age; a substantial reduction in phosphorus; adsorbing coloring materials such as dyes and toxic compounds; and adsorbing detergents and reducing foam.

Equipment normally associated with powdered activated carbon adsorption includes powdered carbon, volumetric and gravimetric feeders, and slurry feeders. A flow diagram of a powdered activated carbon adsorption system is shown below.



IV.5.2.2 Typical Design Criteria

The amount of powdered carbon fed to a system greatly depends on the characteristics of the wastewater and the desired effluent quality; however, powdered carbon will generally be fed at a rate between 50 and 300 mg/L.

IV.5.2.3 Costs

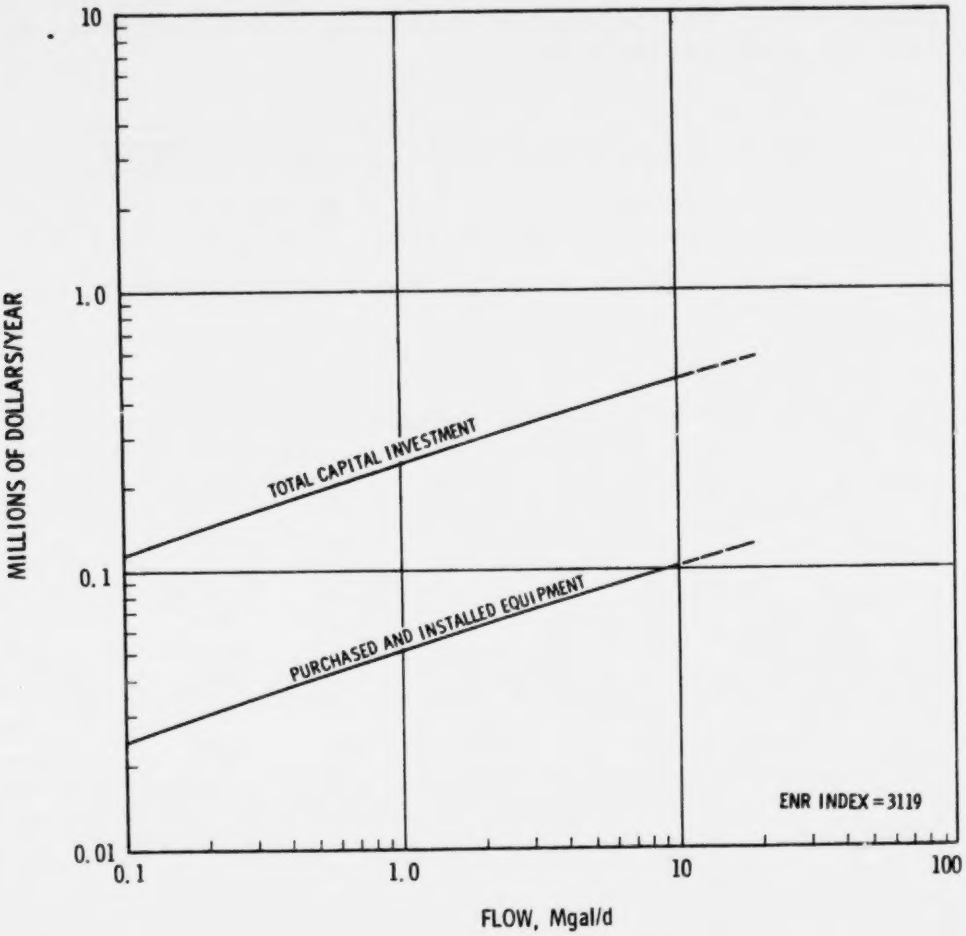
Purchased equipment and installation cost for estimation of the total capital investment includes carbon feeding and storage equipment only. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Carbon dosage	80 mg/L
Carbon feed concentration	1 lb/gal slurry

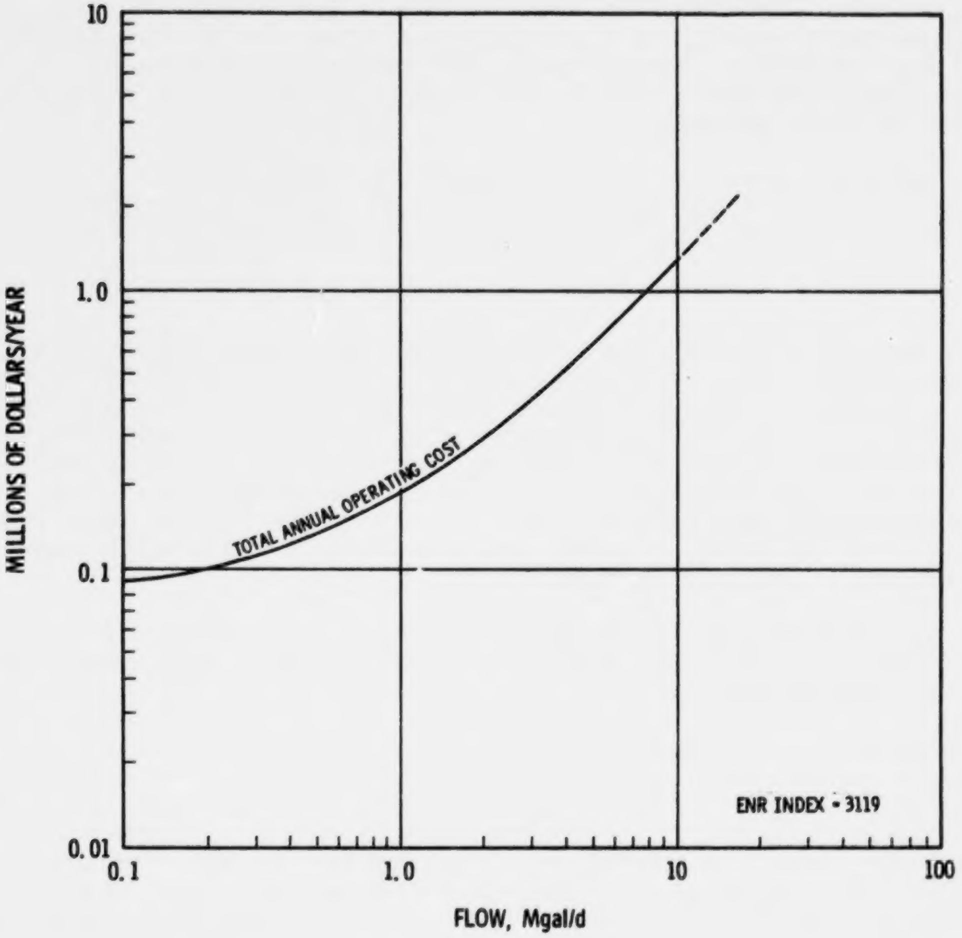
IV.5.2.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.3 CHEMICAL OXIDATION [1]

IV.5.3.1 Description

The chemical oxidation process involves the chemical rather than the biological oxidation of dissolved organics in wastewater.

The processes discussed here are based on chemical oxidation as differentiated from thermal, electrolytic, and biological oxidation. Ozonation, a commonly used chemical method of oxidation for waste treatment, and another oxidation process, chlorination, are discussed elsewhere in this volume. The oxidation reactions discussed here should be distinguished from the higher temperature, and typically pressurized, wet oxidation processes, such as the Zimpro process, which are also discussed in a separate section of this volume.

Oxidation-reduction or "redox" reactions are those in which the oxidation state of at least one reactant is raised while that of another is lowered. In reaction (1) in alkaline solution:

$$2MnO_4^- + CN^- + 2OH^- \rightleftarrows 2MnO_4^{2-} + CNO^- + H_2O \qquad (1)$$

the oxidation state of the cyanide ion is raised from -1 to +1 (the cyanide is oxidized as it combines with an atom of oxygen to form cyanate); the oxidation state of the permanganate decreases from -1 to -2 (permanganate is reduced to manganate). This change in oxidation state implies that an electron was transferred from the cyanide ion to the permanganate. The increase in the positive valence (or decrease in the negative valence) with oxidation takes place simultaneously with reduction in chemically equivalent ratios.

There are many oxidizing agents; however, only a few are convenient to use. Those more commonly used in waste treatment are shown in the following table.

Some oxidations proceed readily to CO_2. In other cases, the oxidation is not carried as far perhaps because of the dosage of the oxidant, the pH of the reaction medium, the oxidation potential of the oxidant, or the formation of stable intermediates. The primary function performed by oxidation in the treatment of hazardous wastes is essentially detoxification. For instance, oxidants are used to convert cyanide to the less toxic cyanate or completely to carbon dioxide and nitrogen. The oxidant itself is reduced. For example, in the potassium permanganate treatment of phenolics, the permanganate is reduced to manganese dioxide. A secondary function is to assure complete precipitation, as in the oxidation of Fe++ to Fe+++ and similar reactions.

WASTE TREATMENT APPLICATIONS OF OXIDATION IDENTIFIED

Oxidant	Waste
Ozone[a]	--
Air (atmospheric oxygen)	Sulfites ($SO_3^=$) Sulfides ($S^=$) Ferrous iron (Fe++) (very slow)
Chlorine gas	Sulfide Mercaptans
Chlorine and gas caustic[b]	Cyanide (CN^-)
Chlorine dioxide	Cyanide Diquat Paraquat
Sodium hypochlorite	Cyanide Lead
Calcium hypochlorite	Cyanide
Potassium permanganate	Cyanide (organic odors) Lead Phenol Diquat Paraquat
Trace quantities only	Organic sulfur compounds Rotenone Formaldehyde
Permanganate	Manganese
Hydrogen peroxide	Phenol Cyanide Sulfur compounds Lead
Nitrous acid	Benzidene

[a] Discussed in another section of this volume.
[b] Alkaline chlorination.

The first step of the chemical oxidation process is the adjustment of the pH of the solution to be treated. In the use of chlorine gas to treat cyanides, for instance, this adjustment is required because acid pH has the effect of producing hydrogen cyanide and/or cyanogen chloride, both of which are poisonous gases. The pH adjutment is done with an appropriate Alkali (e.g., sodium hydroxide). This is followed by the addition of the oxidizing agent. Mixing is provided to contact the oxidizing agent and the waste. Because some heat is often liberated, more concentrated solutions will require cooling. The agent can be in the form of a gas (chlorine gas), a solution (hydrogen peroxide) or perhaps a solid if there is adequate mixing. Reaction times vary but are in the order of seconds and minutes for most of the commercial-scale installations. Additional time is allowed to ensure complete mixing and oxidation.

At this point, additional oxidation may be desired and, as with cyanide destruction, often requires the readjustment of the pH followed by the addition of more oxidant. Once reacted, this final oxidized solution is then generally subjected to some form of treatment to settle or precipitate any insoluble oxidized material, metals, and other residues. A treatment for the removal of what remains of the oxidizing agent (both reacted and unreacted) may be required. A product of potassium permanganate oxidation is manganese dioxide (MnO_2) which is insoluble and can be settled or filtered for removal.

The characteristics of a number of common oxidizing agents are described in the following paragraphs.

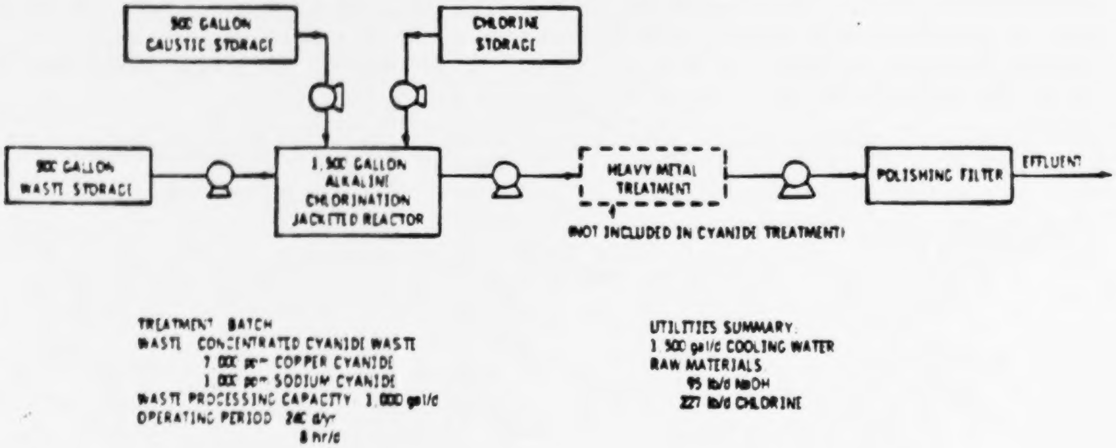
- Potassium Permanganate. Potassium permanganate ($KMnO_4$) has been used for destruction of organic residues in wastewater and in potable water. Its usual reduced form, manganese dioxide (MnO_2), can be removed by filtration. $KMnO_4$ reacts with aldehydes, mercaptans, phenols, and unsaturated acids. It is considered as a relatively powerful oxidizing agent.

- Hydrogen Peroxide. Hydrogen peroxide (H_2O_2) has been used for the separation of metal ions by selective oxidation. In this way it helps remove iron from combined streams by oxidizing the ferrous ion to ferric, which is then precipitated by the addition of the appropriate base. In dilute solution (<30%), the decomposition of hydrogen peroxide is accelerated by the presence of metal ion contaminants. At higher concentrations of hydrogen peroxide, these contaminants can catalyze its violent decomposition. Hydrogen peroxides should be added slowly to the solution with good mixing. This caution relates to other oxidants as well. If the follow-on treatment involves distillation or crystallization, the absence of all unspent peroxides must be confirmed since these techniques tend to concentrate the unused reagent. Hydrogen peroxide has also been used as an "anti-chlor" to remove residual chlorine following chlorination treatment.

- Chromic Acid. Chromium trioxide (CrO_3) commercially called chromic acid, is used as an oxidizing agent in the preparation of organic compounds. It is often regenerated afterward by electrolytic oxidation. In the oxidation of organic compounds, chromic acid in a solution of sulfuric acid is reduced and forms chromium sulfate $[Cr_2(SO_4)_3]$.

Only very simple equipment is required for chemical oxidation. This includes storage vessels for the oxidizing agents and perhaps for the wastes, metering equipment for both streams, and contact vessels with agitators to provide suitable contact of oxidant and waste. Some instrumentation is required to determine the concentration and pH of the water and the degree of completion of the oxidation reaction. The oxidation process may be monitored by an oxidation-reduction potential (ORP) electrode. This electrode is generally a piece of noble metal (often platinum) that is exposed to the reaction medium, and which produces an EMF output that is empirically related to the reaction conditon by revealing the ratio of the oxidized to the reduced constituents.

A flow diagram of chemical oxidation process is shown below.


500 GALLON CAUSTIC STORAGE
CHLORINE STORAGE
500 GALLON WASTE STORAGE
1,500 GALLON ALKALINE CHLORINATION JACKETED REACTOR
HEAVY METAL TREATMENT
(NOT INCLUDED IN CYANIDE TREATMENT)
POLISHING FILTER
EFFLUENT
TREATMENT: BATCH
WASTE: CONCENTRATED CYANIDE WASTE
7,000 ppm COPPER CYANIDE
1,000 ppm SODIUM CYANIDE
WASTE PROCESSING CAPACITY: 1,000 gal/d
OPERATING PERIOD: 240 d/yr
8 hr/d
UTILITIES SUMMARY:
1,500 gal/d COOLING WATER
RAW MATERIALS:
95 lb/d NaOH
227 lb/d CHLORINE

IV.5.3.2 Typical Design Criteria[a]

Design criterion	Value
Tower temperature (top)	185-250°F
Feed	1.4-11.7 lb sulfide/min
Air flow	7.3-12.2 lb air/lb sulfide oxidized
Detention time	1.1-7.9 hr
Pressure (top)	50-79 psia

[a] Design criteria based on sulfide oxidation.

IV.5.3.3 Costs

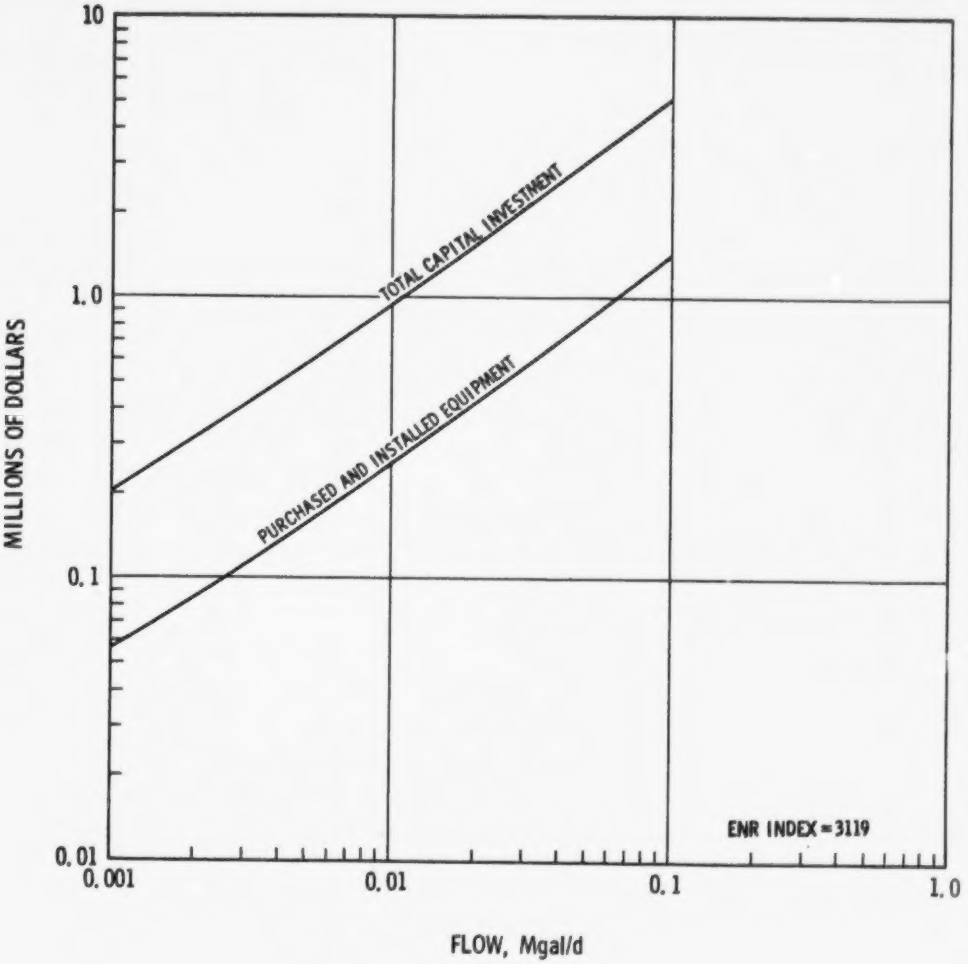
The cost of treatment at individual plant locations and on individual wastes varies greatly. Capital costs depend upon such factors as the type, volume, and composition of the waste; degree of treatment required; treatment process selected; availability of required services; and the specific material to be recovered (i.e., metals, chemicals, or water). For the purpose of cost estimating, data are generated for the chemical oxidation of copper cyanide and sodium cyanide waste from a plating operation using sodium hydroxide and chlorine treatment. A flow diagram was presented in Section IV.5.3.1.

Cost item	Annual quantity	Cost per unit quantity	Annual cost, $	
Direct operating cost				
Labor				
Operating	2,500 man-hr	$16/hr	40,000	
Maintenance			4,700	
				44,700
Chemicals				
Sodium hydroxide	22.9 tons	$160/ton	3,699	
Chlorine	54.5 tons	$160/ton	8,717	
				12,365
Power	35,000 kWh	$0.035/kWh		1,200
Materials				4,900
Total				63,165
Total indirect operating cost				84,500
Total annual operating cost				147,665

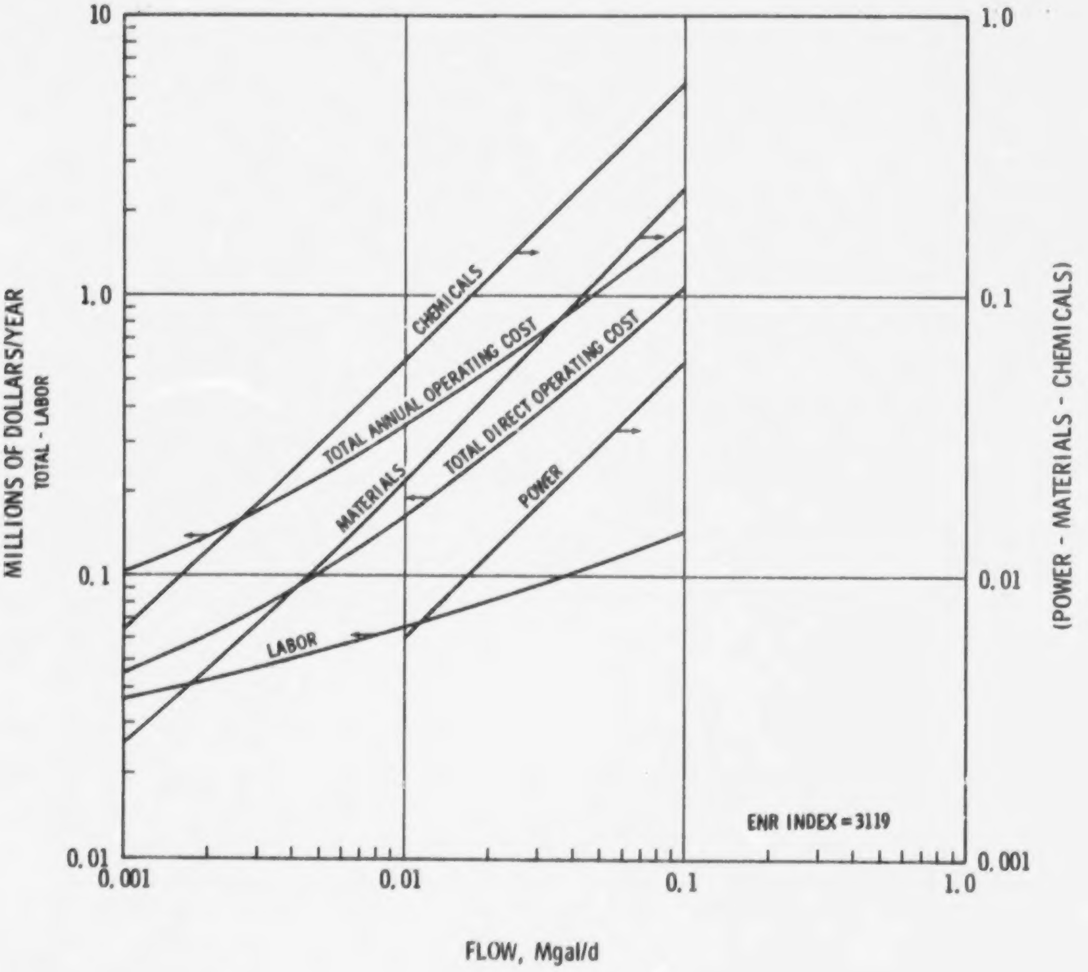
IV.5.3.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 35-1 through 35-19.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.4 AIR STRIPPING [1, 2]

IV.5.4.1 Description

Air stripping of wastewater removes the ammonia nitrogen from the wastewater and discharges it to the air.

Ammonia is quite soluble in water, but this solubility is temperature-dependent. The relationship between temperature and the solubility of ammonia for dilute ammonia solution is expressed by Henry's Law:

$$y = Mx$$

where y = mole fraction NH_3 in the vapor
x = mole fraction NH_3 in the liquid
M = Henry's constant

Henry's constant is a function of temperature. By raising the temperature of the wastewater the vapor pressure of the ammonia is increased, and ammonia removal efficiency increased.

Another factor in ammonia removal efficiency is the pH of the wastewater. A portion of the ammonia dissolved in the water reacts with the water to give the following equilibrium:

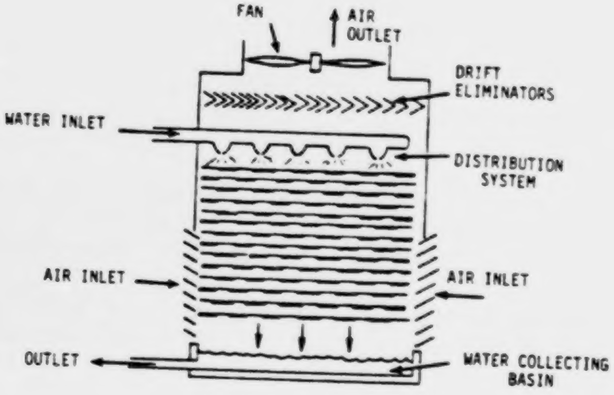
$$NH_3 + H_2O \rightleftharpoons NH_4^+ + OH^- \qquad (1)$$

By increasing the pH (concentration of OH^-), the equilibirum is shifted to the left, reducing the concentration of NH_4^+ and increasing the concentration of free dissolved ammonia.

In air stripping of ammonia from dilute wastewater, the air temperature limits the effectiveness of heating the wastewater. Ammonia removal efficiency is enhanced instead by increasing the pH, usually by the addition of lime. The ammonia-containing wastewater and the lime slurry are fed to a rapid mix tank. Following the rapid mix tank are flocculators and a settling basin, where calcium phosphate precipitates and recirculated calcium carbonate settles out. The clarified, lime-treated, wastewater is pumped to the top of packed towers. In each tower, fans draw air up through the tower countercurrent to the falling wastewater. The "packing" in the tower may be slats or a series of bundles of pipe with the pipe sections spaced 2 to 3 inches on center. The pipe sections are horizontal, and the direction of each row alternates.

After the wastewater has been air stripped of ammonia, it flows into the recarbonation basin where compressed carbon dioxide rich gas from the lime recalcining furnace is bubbled through it to precipitate calcium carbonate. Some of the calcium carbonate sludge is returned to the rapid mix tank to enhance flocculation while the remainder of the calcium carbonate sludge and the phosphate sludge from the settling basins are sent to centrifuges. The sludges can be fractionally centrifuged to yield two dewatered sludges, one rich in calcium carbonate and one containing phosphate.

Equipment normally associated with the air stripping process includes the stripping tower, which closely resembles a conventional cooling tower. A flow diagram of the air stripping process is shown below.


FAN
AIR OUTLET
DRIFT ELIMINATORS
WATER INLET
DISTRIBUTION SYSTEM
AIR INLET
AIR INLET
OUTLET
WATER COLLECTING BASIN

COUNTERCURRENT TOWER

IV.5.4.2 Typical Design Criteria

Typical design criteria for an air stripping system are listed below.

Design criterion	Value
Wastewater loading	1 - 2 gpm/ft²
Stripping air flow rate	300 - 500 ft³/gal
Packing depth	20 - 25 ft
pH of wastewater	10.8 - 11.5
Air pressure drop	0.015 - 0.019 in. of water/ft
Packing material	Plastic or wood
Packing spacing	Approximately 2 in. horizontal and vertical
Provisions required	Uniform water distribution, and scale removal and cleanup
Land requirement	Small

IV.5.4.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes tower, piping, and pump. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Pump total discharge head	50 ft
Loading	1 gpm/ft^3
Air flow	400 ft^3/gal
Ammonia concentration	18 mg/L influent; 3 mg/L effluent
pH	11 - 11.5
Tower height	20 ft
Indicate line dosage	Typical

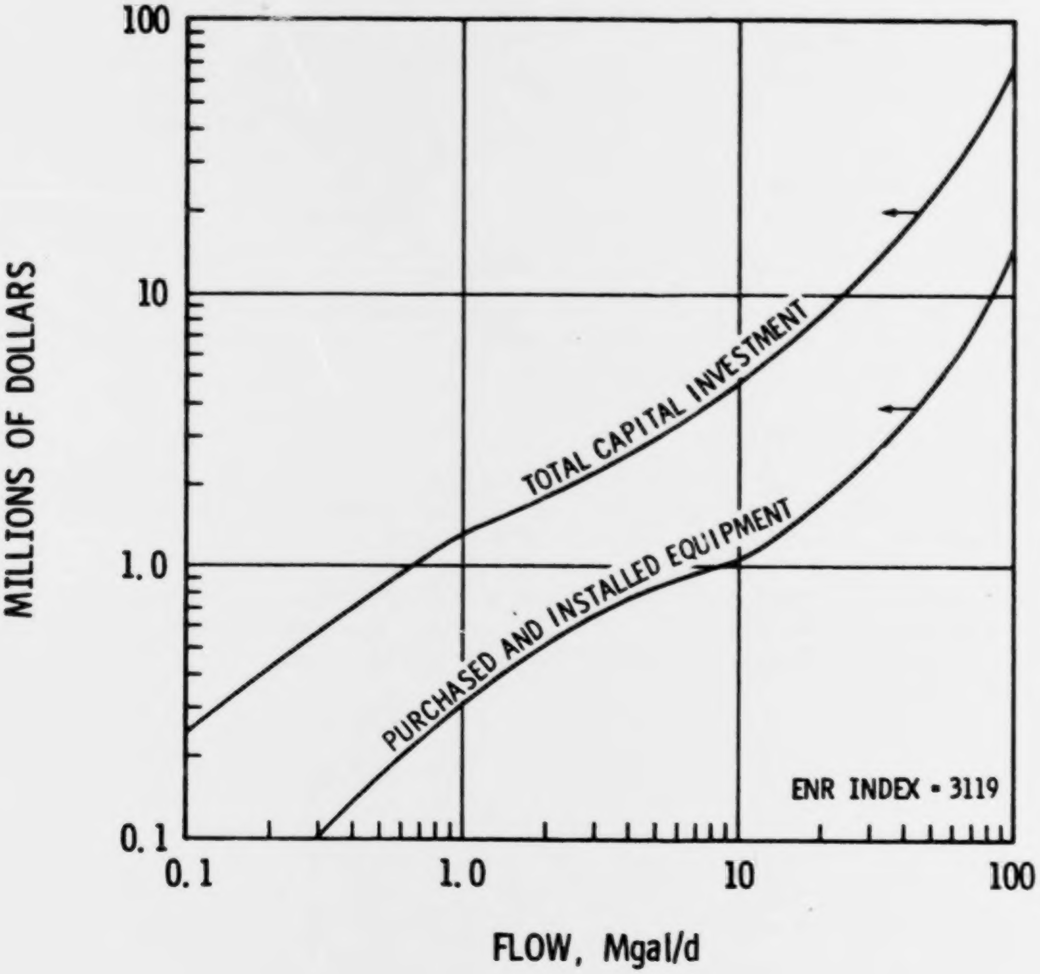
The tower is packed with 1/2-inch diameter Schedule-80 PVC pipes. The distances between these pipes is 3 inches in horizontal rows and 2 inches in vertical columns.

IV.5.4.4 References

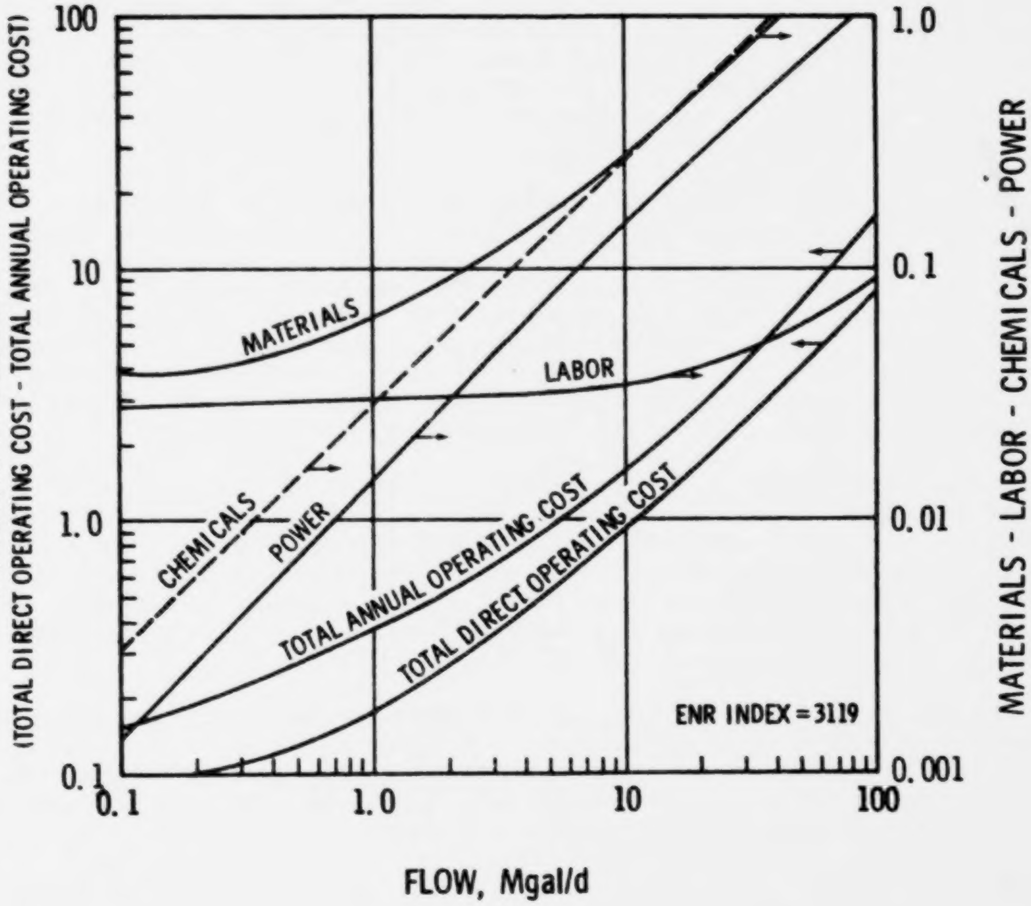
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

2. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, DC, November 1976. pp. 41-1 through 41-15.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.5 NITRIFICATION [1,2]

IV.5.5.1 Description

Nitrification is used for the biological oxidation of ammonia to nitrates and nitrites. This process is called single-stage nitrification, because ammonia and carbonaceous materials are oxidized in the same aeration unit. As in any aerobic biological process, carbonaceous materials are oxidized by heterotrophic aerobes. In addition, a special group of autotrophic aerobic organisms called nitrifiers oxidize ammonia in two stages: Nitrosomonas baceteria convert ammonia to nitrate, and Nitrobacter bacteria convert nitrite to nitrate. The optimal conditions for nitrification, in general, include a temperature of about 30°C, pH of about 7.2 to 8.5, F/M of about .10 to .20, relatively long aeration detention times (as nitrifiers have a lower growth rate than other aerobes), and sludge retention times of about 25 to 20 days, depending upon temperature.

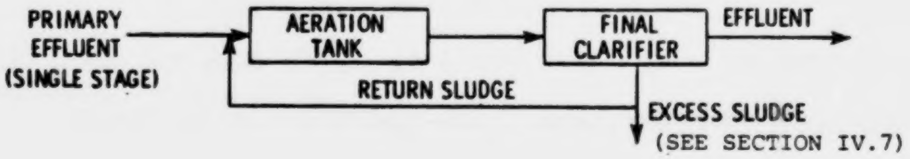
The degree of nitrification depends on a number of factors: sludge retention time (SRT) as a function of temperature, mixed liquor DO concentration, pH, and the absence of inhibitors. If the sludge is wasted at a rate that is too high, the nitrifiers will be eliminated from the system. The aeration system is designed to provide the additional oxygen. Alkalinity is destroyed during nitrification (7.15 mg A' N as $CaCO_3$/mg NO_3-N) and thus alkalinity addition may be required for high TNN (Total Niedahl Nitrogen) low buffer capacity wastewaters.

The conventional and high-rate modifications of the activated sludge process do not provide the necessary hydraulic and sludge detention times; in addition, the F/M ratio is higher. As a result, single-stage nitrification is not normally achieved in these configurations, although they effect a small reduction (about 20 percent) in ammonia due to incorporation into microbial cells.

Any low-rate modification of the activated sludge process such as extended aeration and the oxidation ditch can be used. In addition, the use of powdered activated carbon has the potential of providing consistent ammonia removal, although its application is in a state of infancy and costs are relatively high.

Another modification involves the use of separate stage nitrification. In this modification, carbonaceous oxidation and nitrogenous oxidation are treated in two separate aeration basin and clarifier systems. This provides some degree of production to the nitrification system when inhibiting organics present in the raw wastewater removed in the first carbonaceous stage.

Equipment normally associated with a nitrifying system includes air diffusers, aeration tanks, clarifier equipment, controls, and instrumentation. A flow diagram of the nitrification system is shown below.


PRIMARY EFFLUENT (SINGLE STAGE)
AERATION TANK
FINAL CLARIFIER
EFFLUENT
RETURN SLUDGE
EXCESS SLUDGE
(SEE SECTION IV.7)

IV.5.5.2 Typical Design Criteria

Typical design criteria are tabularized below.

Design criterion	Value
Type of reactor	Plug-flow
Aeration system	Oxygen or air
Mean cell residence time	10 - 20 d
MLVSS[a]	1,000 - 2,000 mg/L
pH	7.2 - 8.5

[a] Mixed liquor volatile suspended solids.

IV.5.5.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes nitrification tanks, aeration devices, clarifiers, and sludge recycle and waste pumps. Facilities for pH adjustment are not included. The system is designed to follow a high-rate activated sludge system. The following operating characteristics were assumed for cost estimation:

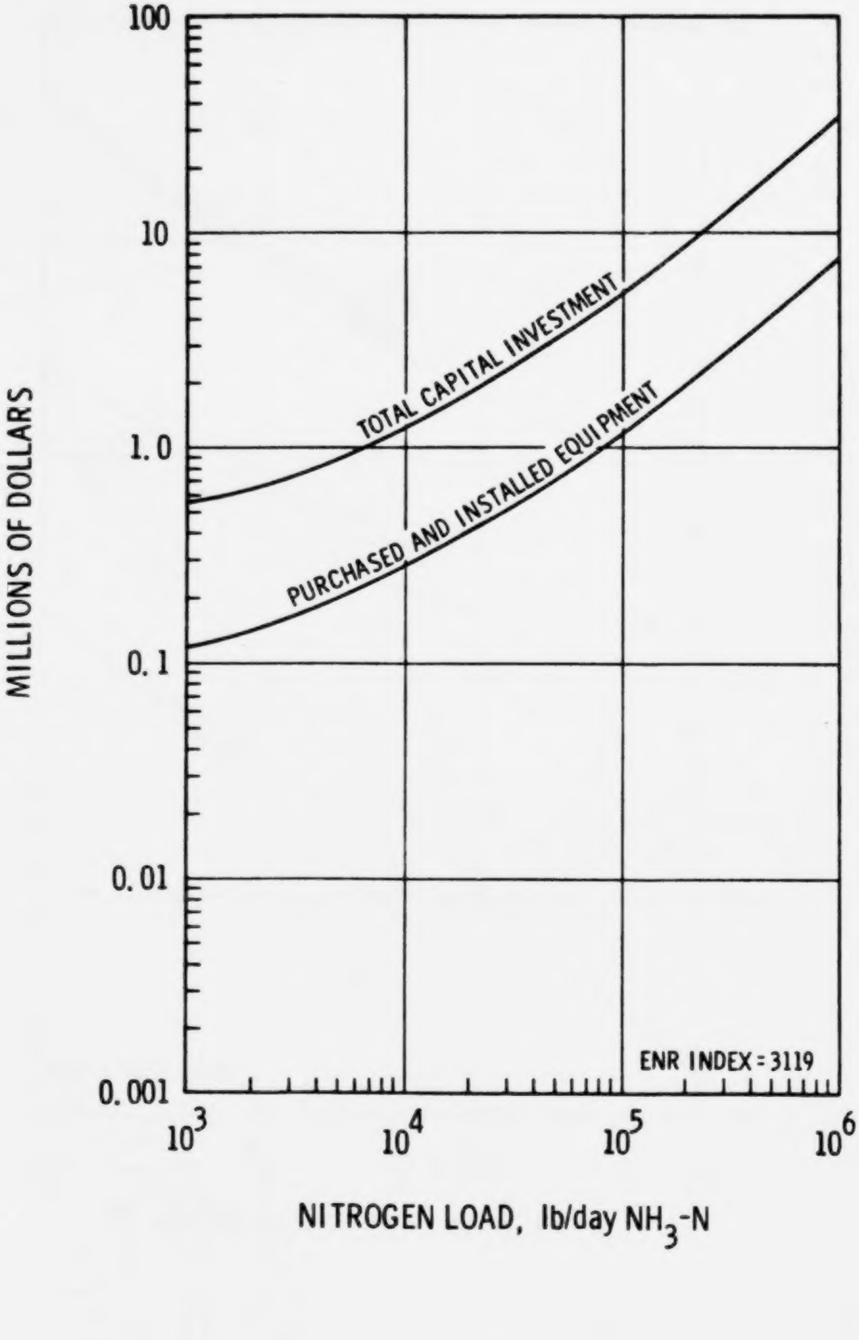
Operating characteristic	Assumed value
Nitrogen loading	1800 mg/L NH_3-N
Sludge age	Design parameter
Oxygen requirement	1.5 lb O_2/lb BOD_5 removed; plus 4.6 lb O_2/lb NH_4-N (oxidized)
Final clarifier overflow rate	600 gpd/ft^2 (30 lb/ft^2/d)
Sludge return pumps sized for 100% recycle and operated at 50% recycle.	

IV.5.5.4 References

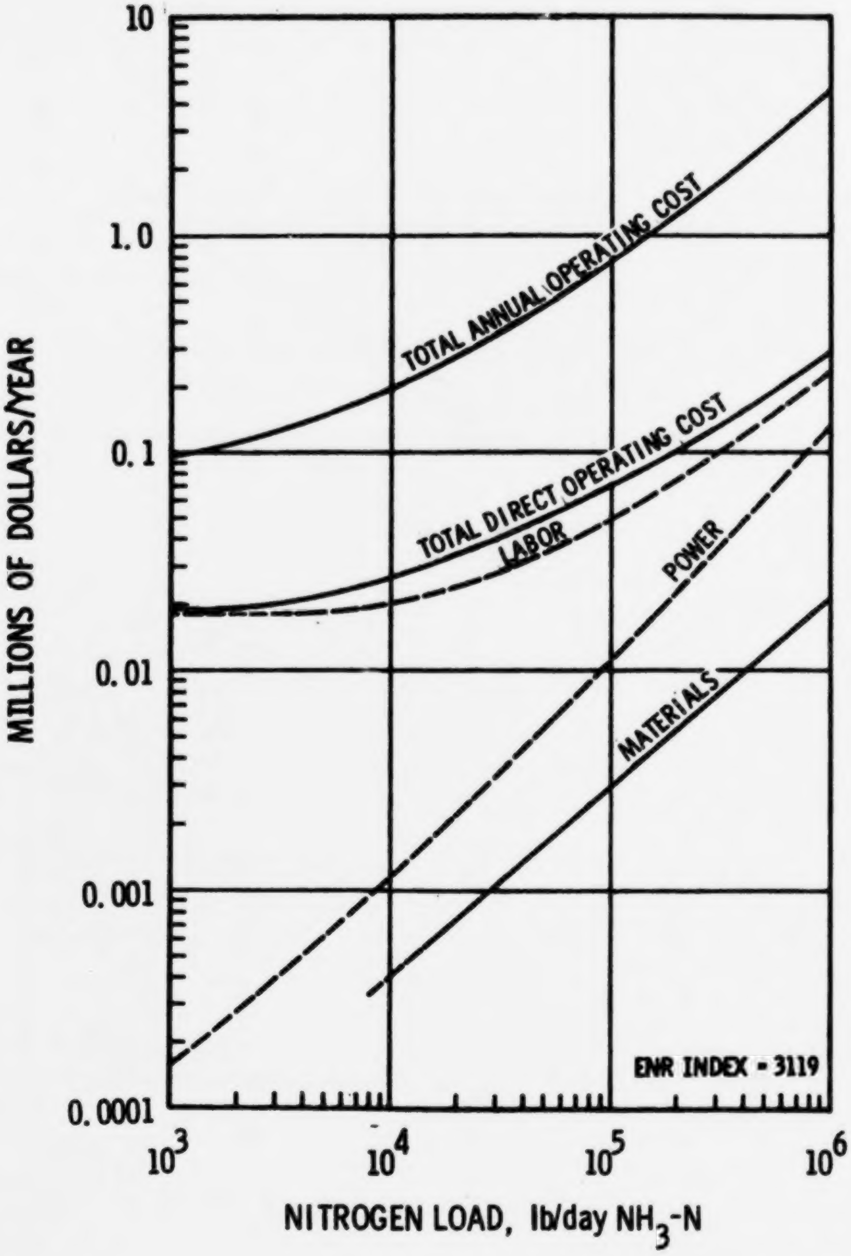
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

2. Metcalf and Eddy. Wastewater Engineering: Collection, Treatment, Disposal. McGraw-Hill Book Co., New York, New York, 1972. pp. 662-667.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



Date: 6/23/80

IV.5.6 DENITRIFICATION [1]

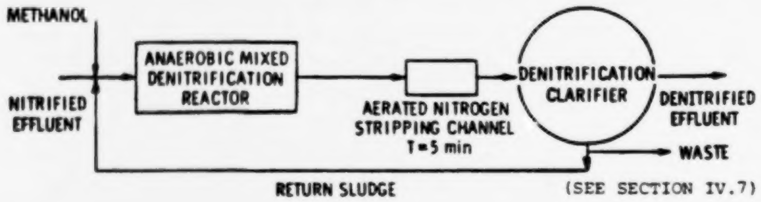
IV.5.6.1 Description

Denitrification is used for the reduction of nitrates and nitrites to nitrogen gas through the action of facultative heterotrophic bacteria. In suspended-growth, separate-stage denitrification processes, nitrified wastewater containing primarily nitrates is passed through a mixed anaerobic vessel containing denitrifying bacteria. Because the nitrified feedwater contains very little carbonaceous material, a supplemental source of carbon is required to maintain the denitrifying biomass. This supplemental energy is provided by feeding methanol to the biological reactor along with the nitrified wastewater. Mixing in the anaerobic denitrification reaction vessel may be accomplished using low speed paddles analogous to standard flocculation equipment. Following the reactor, the denitrified effluent is aerated for a short period (5 to 10 min) to strip out gaseous nitrogen formed in the previous step that might otherwise inhibit sludge settling. Clarification follows the stripping step with the collected sludge being either returned to the head end of the denitrification system or wasted.

Common modifications include the use of alternate energy sources such as sugars, acetic acid, ethanol or other compounds. Nitrogen-deficient materials, such as brewery wastewater, may also be used. An intermediate aeration step for stabilization (about 50 min) between the denitrification reactor and the stripping step may be used to guard against carryover of carbonaceous materials. The denitrification reactor may be covered but not air tight to assure anaerobic conditions by minimizing surface reaeration.

Additional modification include the use of anaerobic-aerobic zones in a nitrifying reactor to promote dentrification using the raw waste carbonaceous BOD such as used in the Caroucel Process. This may also be accomplished by effluent recycle from the nitrifying reactor to the lead of the aeration task which is maintained in an anaerobic condition as practiced in the Bardenpho process. For high rate dentrification requiring minimal land area, fluidized bed processes have recently been developed.

Equipment normally associated with the denitrification process includes clarifier equipment, controls, air diffuser, aeration tanks, instrumentation, chemical feed equipment, and flocculators. A flow diagram of a denitrification system is shown below.


METHANOL
NITRIFIED EFFLUENT
ANAEROBIC MIXED DENITRIFICATION REACTOR
AERATED NITROGEN STRIPPING CHANNEL T=5 min
DENITRIFICATION CLARIFIER
DENITRIFIED EFFLUENT
WASTE
RETURN SLUDGE
(SEE SECTION IV.7)

IV.5.6.2 Typical Design Criteria

Typical design criteria for denitrification are listed below.

Criteria	Value
Flow scheme	Plug flow (preferable)
Optimum pH	6.5 - 7.5
Mixed liquor volatile suspended solids	1,000 - 3,000 mg/L
Mixer power requirement	0.25 - 0.5 hp/1,000 ft^3
Clarifer surface loading rate	400 - 600 gpd/ft^2
Solids loading	20 - 30 lb/d/ft^2
Return sludge rate	50 - 100%
Sludge generation	0.2 lb/lb CH_3OH or 0.7 lb/lb NO_3-N reduced
Hydraulic detention time	0.2 - 2 hr
Mean cell residence time	1 - 5 d

IV.5.6.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes denitrification tanks (uncovered), mixers, methanol feed, clarifiers, and sludge recycle and waste pumps. The reaeration facility is not included. The operating characteristics shown below were assumed for cost estimation.

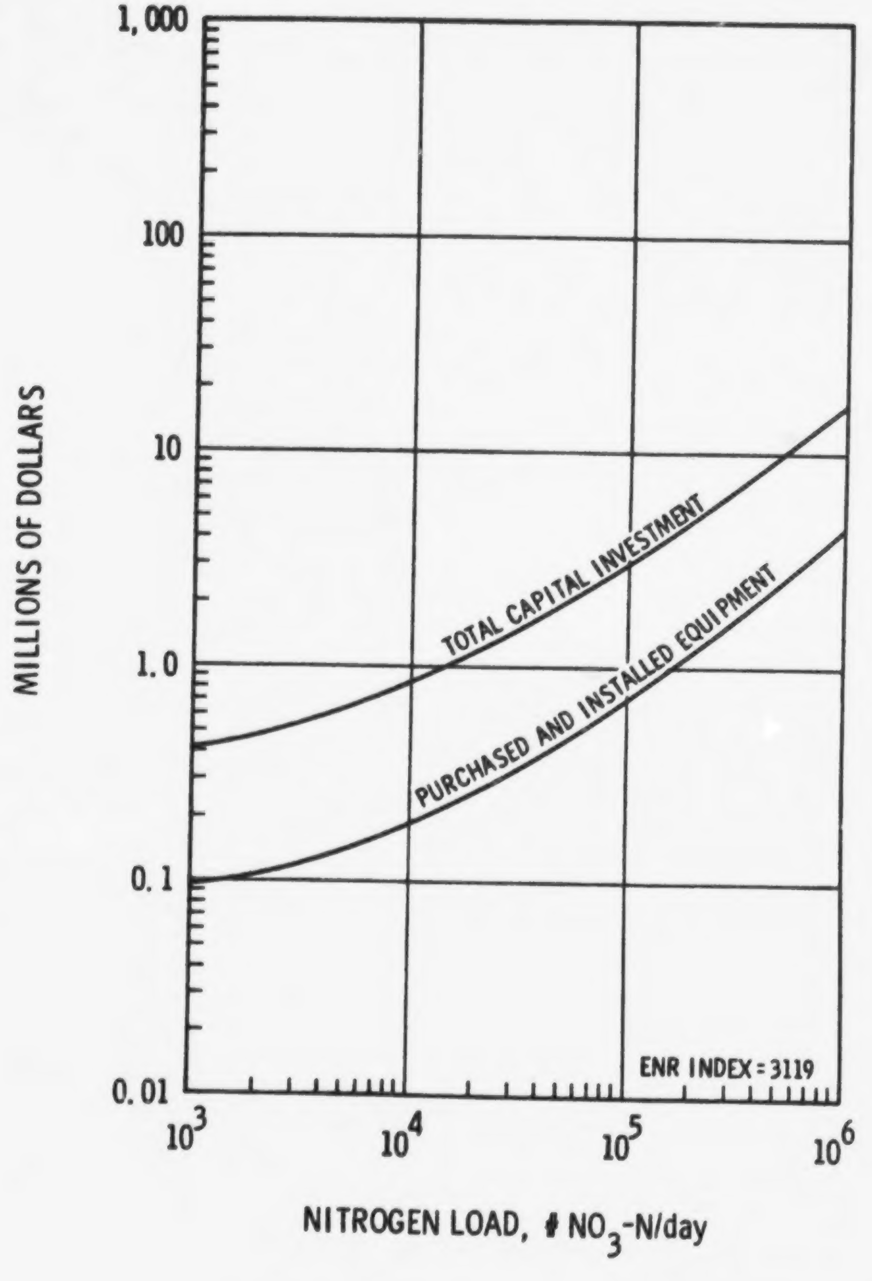
Operating characteristic	Assumed value
Detention time	2 hr in denitrification tank
MLVSS[a]	2,000 mg/L
Final clarifier overflow rate	600 gpd/ft^2
Methanol requirement	3 lb methanol/lb nitrate nitrogen removed
Methanol storage	30 d supply with a minimum tank size of 500 gal
Denitrification recycle pumps	Sized for 100% recycle but operated at 50% recycle
NO_3-N concentration	18 mg/L influent

[a] Mixed liquor volatile suspended solids.

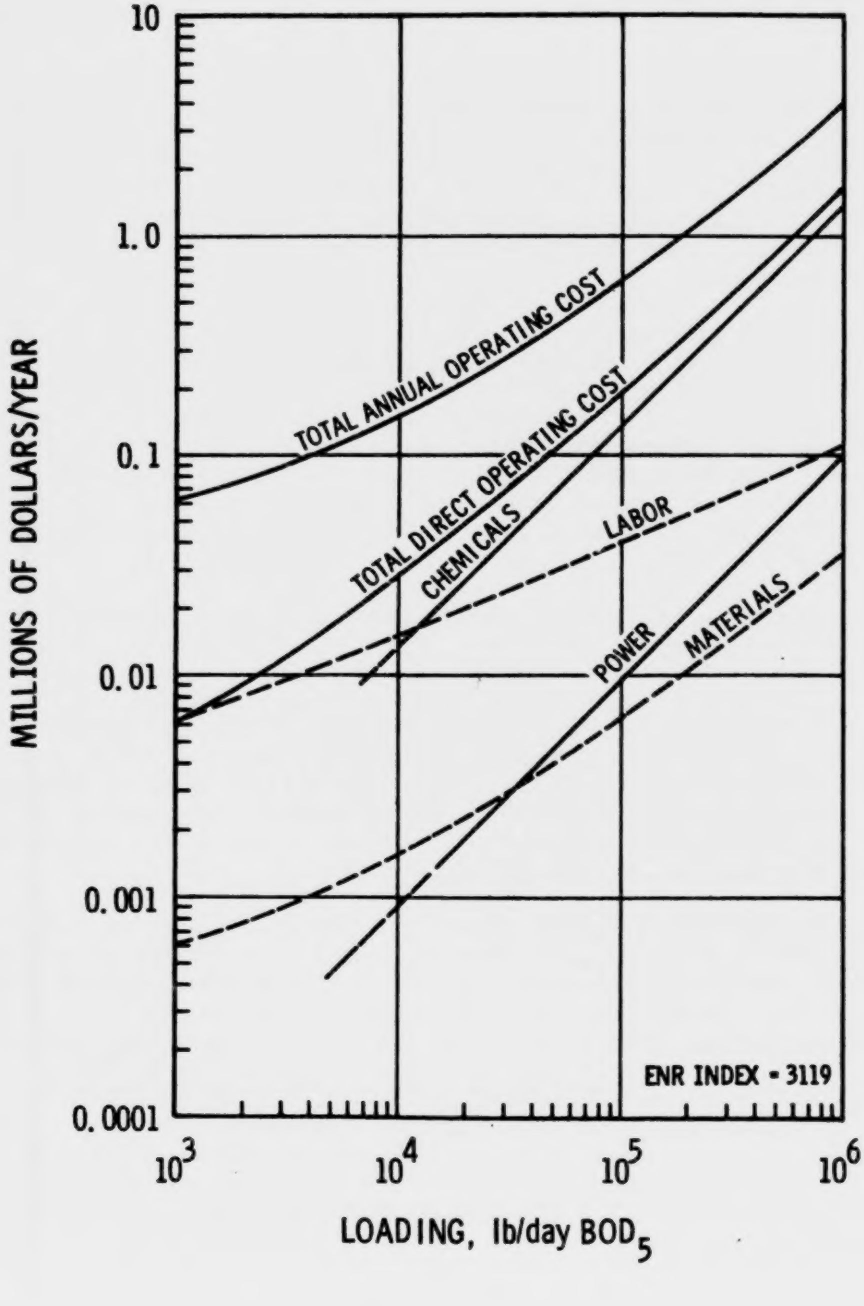
IV.5.6.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.7 ION EXCHANGE [1]

IV.5.7.1 Description

In simplest terms, ion exchange may be thought of as the reversible interchange of ions between an insoluble, solid salt (the "ion exchanger") and a solution of electrolyte in contact with that solid.

In the customary mode of usage, the ion exchanger is contacted with the solution containing the ion to be removed until the active sites in the exchanger are partially or completely used up ("exhausted") by that ion. The exchanger is then contacted with a sufficiently concentrated solution of the ion originally associated with it for conversion ("regeneration") back to its original form.

The ion exchange process works well with cations (including, of course, the hydrogen ion) and anions, both inorganic and organic. However, the organic species frequently interact with the exchangers (particularly the organic resins) via both absorption and ion exchange reactions, often necessitating the use of extremely high regenerant concentrations and/or the use of organic solvents to remove the organics. Consequently, most of the applications of ion exchange of interest have involved inorganic species.

There are a variety of different cation and anion exchangers that form salts of more or less different stabilities with a particular ion. Thus, knowledgeable choice of a particular ion exchange material will often allow selective separation of one ion in solution from another, and afford selective removal of an undesirable ion from a number of innocuous ones. As a general rule, ions with a higher charge will form more stable salts with the exchanger than those with a lower charge, and hence polyvalent species can frequently be selectively removed from a solution of monovalent ones.

In carrying out ion exchange reactions in a column or bed operation (as opposed to a stirred batch operation which is occasionally used in chemical processing), there are four operations carried out in a complete cycle: service (exhaustion), backwash, regeneration, and rinse. The service and regeneration steps have been described above. The backwash step is one in which the bed is washed (generally in water) in reverse direction to the service cycle in order to expand and resettle the resin bed. This step eliminates channeling which might have occurred during service and removes fines or other material that may be clogging the bed. The rinse step removes the excess regeneration solution prior to the next service step.

There are three principal operating modes in use today: cocurrent fixed-bed, countercurrent fixed-bed, and continuous countercurrent. A comparison summary is presented in the following table.

COMPARISON OF ION EXCHANGE OPERATING MODES

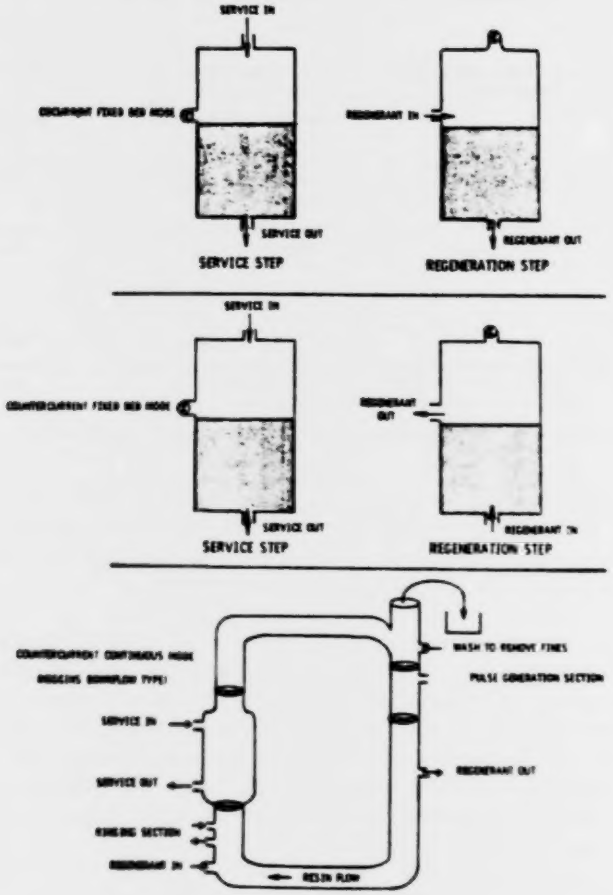
Operating characteristic	Concurrent fixed bed	Countercurrent fixed bed	Countercurrent continuous
Capacity for high feed flow and concentration	Least	Middle	Highest
Effluent quality	Fluctuates with bed exhaustion	High with minor fluctuations	High
Regenerant and rinse requirements	Highest	Somewhat less than concurrent	Least, yields most concentration regenerant waste
Equipment complexity	Simplest; can use manual operation	More complex; automatic controls for regeneration	Most complex; completely automated
Equipment for continuous operation	Multiple beds, single regeneration equipment	Multiple beds, single regeneration equipment	Provides continuous service
Relative costs (per unit volume)			
Investment	Least	Middle	Highest
Materials/labor	Highest chemicals and labor needs, highest resin inventory	Less chemicals, water and labor than cocurrent	Least chemicals, and and labor needs, lowest resin inventory

Most ion exchange installations in use today are of the fixed-bed type, with countercurrent operation coming more into favor, especially for removal (polishing) of traces of hazardous species from the stream prior to reuse or discharge.

In order to minimize regeneration chemical requirements (i.e., to make most efficient use of regenerant), many fixed-bed installations use a technique termed "staged," or "proportional," regeneration. The first part of the regeneration solution to exit from the ion exchange bed is the most enriched in the component being removed; the concentration of that component decreases in succeeding portions of the exiting regeneration solution. In staged regeneration, the solution is divided (generally in separate tanks) into two or more portions. The first portion through the bed is "discarded" (i.e., sent for subsequent treatment), while the second and succeeding portions (less rich in the species being removed) are retained. On the next regeneration cycle, the second portion from the preceding cycle is passed through the bed first (and then "discarded"), followed by the succeeding portions, the last of which is a portion of fresh regenerant. In this way, regenerant utilization can be maximized.

Fixed-bed ion exchange operations are straightforward systems, requiring a cylindrical ion exchange bed, tanks for solution storage, and pumps. The choice of materials is governed by the chemical environment. Continuous ion exchange systems are much more complex, requiring solids handling equipment and more intricate control systems. Apparently only one company (Chemical Separations Corp.) has been truly successful in the design and fabrication of continuous ion exchange systems, and it should be consulted if the use of such a system is contemplated.

A flow diagram of ion exchange systems is shown below.


SERVICE IN
COCURRENT FIXED BED MODE
SERVICE OUT
SERVICE STEP
REGENERANT IN
REGENERANT OUT
REGENERATION STEP
SERVICE IN
COUNTERCURRENT FIXED BED MODE
SERVICE OUT
SERVICE STEP
REGENERANT OUT
REGENERANT IN
REGENERATION STEP
COUNTERCURRENT CONTINUOUS MODE
(HIGGINS DOWNFLOW TYPE)
SERVICE IN
SERVICE OUT
RINSING SECTION
REGENERANT IN
RESIN FLOW
WASH TO REMOVE FINES
PULSE GENERATION SECTION
REGENERANT OUT

IV.5.7.2 Typical Design Criteria

Design criterion	Value
Clinoptilolite size	20 x 50 mesh
Bed height	4 to 6 ft
Wastewater suspended solids	35 mg/L max
Wastewater loading rate	7.5 to 20 bed volume/h
Pressure drop	8.4 in of water/ft
Cycle time	100 to 150 bed volumes for one 6 ft bed; 200 to 250 bed volumes for two 6 ft beds in series

[a] Ion exchange for ammonia removal.

IV.5.7.3 Costs

Ion exchange is frequently incorporated into systems with other treatment processes, such as reverse osmosis. In most cases, separate cost data are not available solely for ion exchange. In addition, many practitioners are reluctant to divulge treatment cost data. Additionally, costs vary widely and depend primarily upon the stream size and composition. Care should be exercised in comparing costs for different processes because the product streams are rarely the same.

For the purpose of cost estimating, data are generated for a dilute mixed acid waste stream (80,000 gpd) from a metal finishing operation. Treatment is based on the following wastewater characteristics shown below. Product streams were clean water, metal cations, chromium, and cyanide.

Pollutant	Concentration	
	Feed stream to process	Effluent
Zinc	15 mg/L	<1 mg/L
Copper	0.5 mg/L	<0.3 mg/L
Cyanide (total)	19 mg/L	<0.3 mg/L
Chromium	22 mg/L (mostly Cr^{+6})	<0.3 mg/L
pH	∿10	6 - 8.5

Design basis for the above operation included a three-bed system, in duplicate for regeneration; a 24 hr, 350 d/yr operation; and a 24-hr loading time.

Fixed capital costs based on an ENR index of 3119 are estimated to be $540,000. Similarly, total capital cost is estimated to be $570,000. Estimation of annual operating cost is presented in the following table.

Cost item	Annual quantity	Cost per unit quantity	Annual cost,[a] $	
Direct operating cost				
Labor	2,200/man-hr	$16/hr		35,200
Chemicals				
NaOH (70%)	175 tons	$160/ton	28,000	
H_2SO_4 (98%)	48 tons	$55/ton	2,600	
Resin replacement	20%		3,200	
				33,800
Power (electricity)	75,000 kWh	$0.035/kWh		2,600
Total				71,600
Total indirect operating cost				103,600
Total annual operating cost				175,200

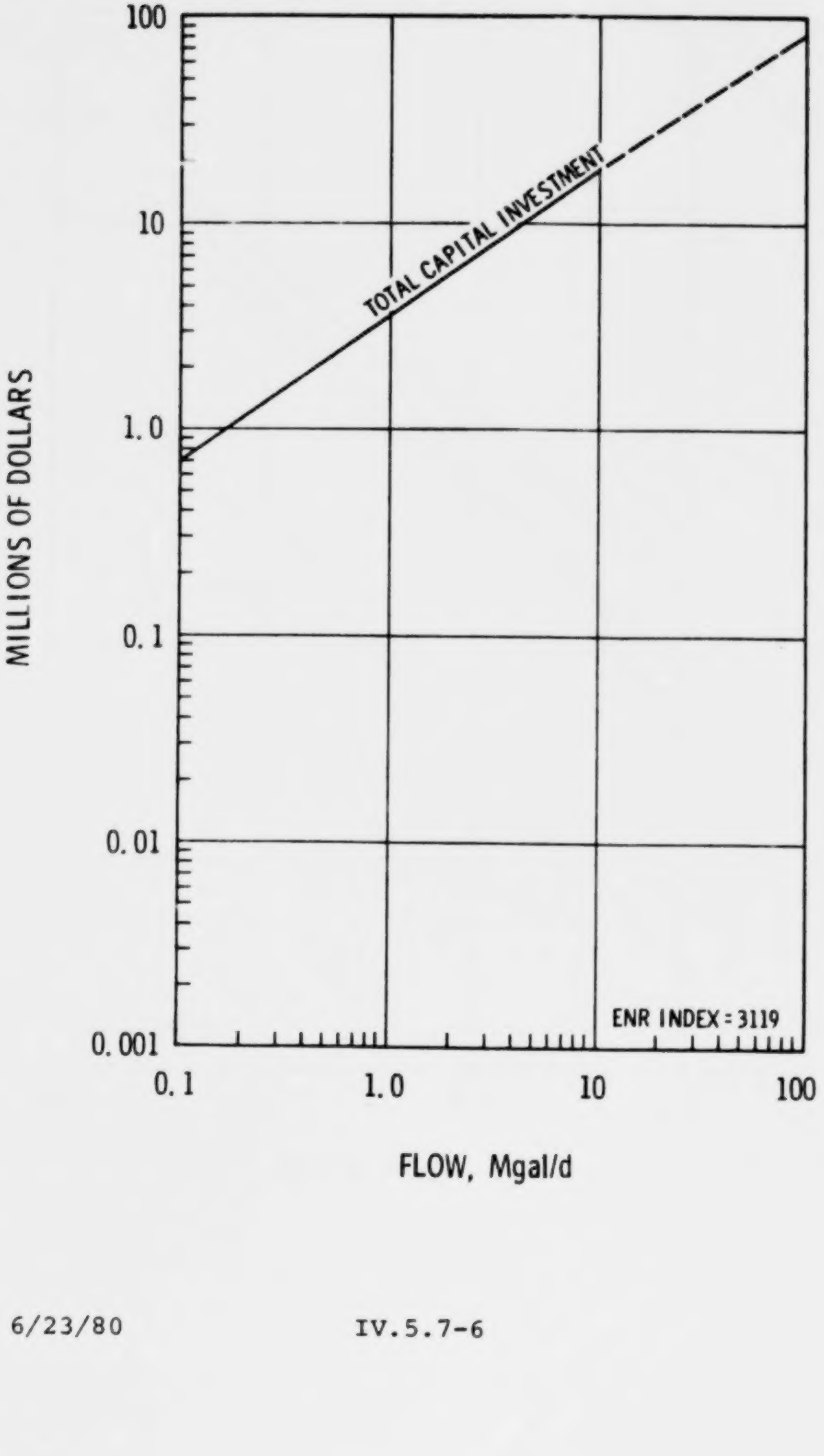
Cost curves were developed for total capital investment and annual operating cost using the cost data shown on the previous page and the following exponential scaling factors, which were used to determine costs at varying flowrates:

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power and chemicals	1.0

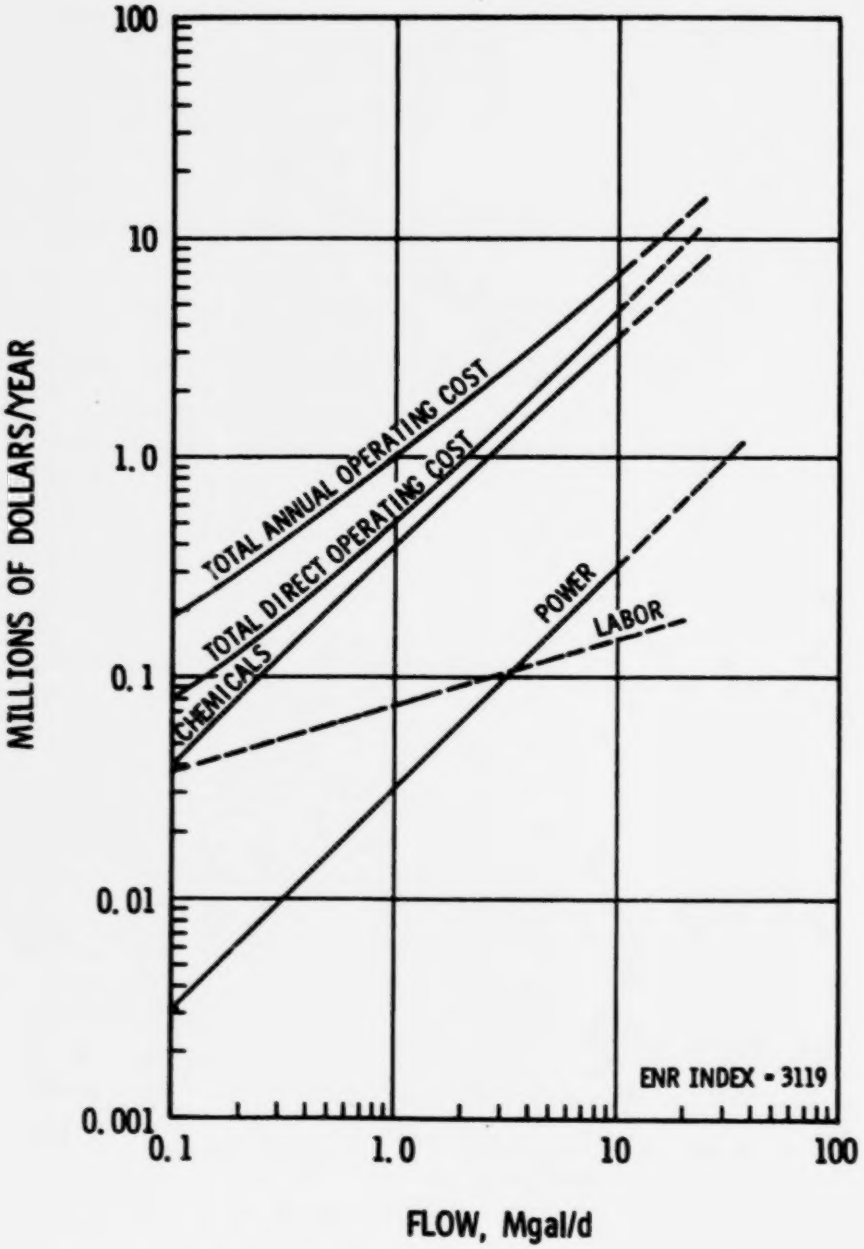
IV.5.7.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 30-1 through 30-26.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.8 POLYMERIC (RESIN) ADSORPTION [1]

IV.5.8.1 Description

Adsorption on synthetic resins is considered here primarily as a process for the removal of organic chemicals from liquid waste streams. Waste treatment by resin adsorption involves two basic steps: (1) contacting the liquid waste stream with the resins and allowing the resins to adsorb the solutes from the solution; and, (2) subsequently regenerating the resins by removing the adsorbed chemicals, often effected by simply washing with the proper solvent.

The chemical nature of the various commercially available resins can be quite different; perhaps the most important variable in this respect is the degree of their hydrophilicity. The adsorption of a nonpolar molecule on to a hydrophobic resin (e.g., a styrene-divinyl benzene based resin) results primarily from effect of van der Waal's forces. In other cases, other types of interactions such as dipole-dipole interaction and hydrogen bonding are also important. In a few cases, an ion-exchange mechanism may be involved; this is thought to be true, for example, in the adsorption of alkylbenzene sulfonates from aqueous solution onto weakly basic resins; e.g., a phenol-formaldehyde-amine based resin.

Resin adsorbents are used in much the same way as granular carbon. Commonly, a typical system for treating low volume waste streams will consist of two fixed beds of resin. One bed will be on stream for adsorption while the second is being regenerated. In cases where the adsorption time is very much longer than the regeneration time (e.g., when solute concentrations are very low), one resin bed plus a hold-up storage tank could suffice.

The adsorption bed is usually fed downflow at flow rates in the range of 0.25 to 2 gpm per cubic foot of resin; this is equivalent to 2 to 16 bed volumes/hr, and thus contact times are in the range of 3 to 30 minutes. Linear flow rates are in the range of 1 to 10 gpm/ft^2. Adsorption is stopped when the bed is fully loaded and/or the concentration in the effluent rises above a certain level.

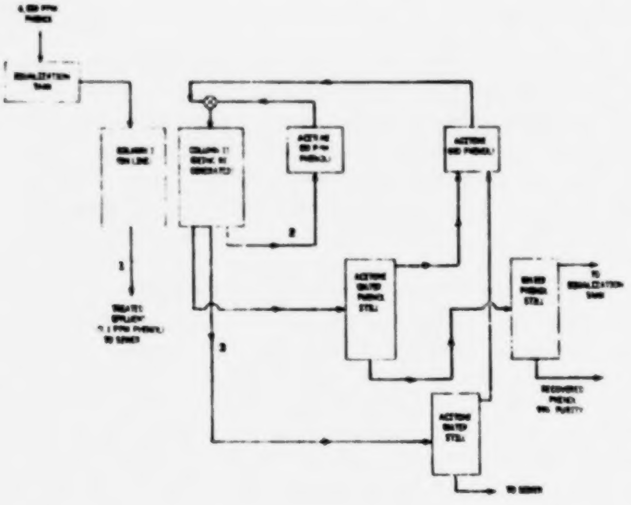
Regeneration of the resin bed is performed *in situ*; basic, acidic, and salt solutions or regenerable nonaqueous solvents are most commonly used. Basic solutions may be used for the removal of weakly acidic solutes, and acidic solutions for the removal of weakly basic solutes; hot water or steam could be used for volatile solutes; methanol and acetone are often used for the removal of nonionic organic solutes. A prerinse and/or a post-rinse with water will be required in some cases. As a rule, about three bed volumes of regenerant will be required for resin regeneration; as little as one-and-a-half bed volumes may suffice in certain applications.

Solvent regeneration will be required unless (1) the solute-laden solvent can be used as a feed stream in some industrial process at the plant, or (2) the cost of the solvent is low enough so that it may be disposed of after a single use. Solvent recovery, usually by distillation, is thus most common when organic solvents are used. Distillation will allow solute recovery for reuse if such is desired.

Resin lifetimes may vary considerably depending on the nature of the feed and regenerant streams. Regeneration with caustic is estimated to cause a loss of 0.1 to 1% of the resin per cycle; replacement of resins at such installations may be necessary every two to five years. Regeneration with hot water, steam, or organic solvent should not affect the resins, and, in this case, lifetimes will be limited by slow fouling or oxidation resulting in a loss of capacity; actual experience indicates that lifetimes of more than five years are obtainable.

Equipment for resin adsorption systems is relatively simple. The system will generally consist of two or more steel tanks (stainless or rubber-lined) with associated piping, pumps, and (perhaps) influent hold-up tank. Regeneration takes place in the same tanks, and thus the extra equipment needed for regeneration will consist only of such items as solvent storage tanks, associated solvent piping and pumps, and solvent (and perhaps solute) recovery equipment; e.g., a still. Up to three stills may be required in some systems.

A flow diagram of a polymeric adsorption system is shown below.



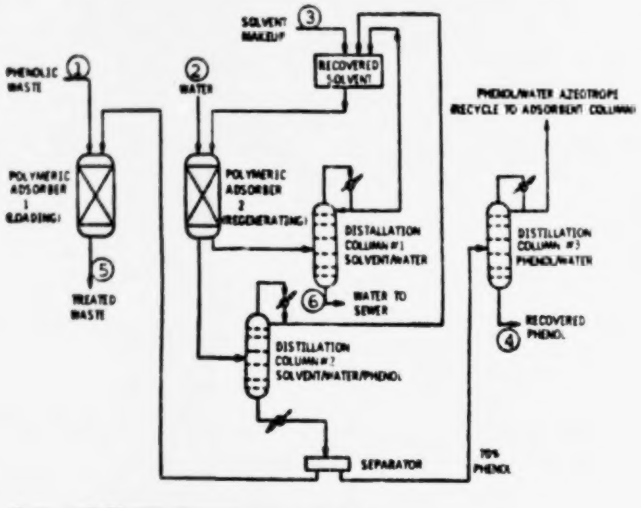
IV.5.8.2 Typical Design Criteria

Design criterion	Value
Flow rates (downflow)	0.25 to 2 gpm/foot of resin
Contact time	3 to 30 min
Suspended solid	<50 ppm
Surface areas of resin adsorbents	800-1,200 m^2/g

IV.5.8.3 Costs

Costs of wastewater treatment by polymeric adsorption can vary widely depending on the nature of waste being treated and the system design. As with carbon adsorption systems, two primary variables in the process economics are: (1) the polymer exhaustion rate (i.e., the amount of polymer required to treat a given volume of liquid), and (2) the superficial liquid retention time (i.e., the time that the liquid takes to fill the volume of the polymer bed). For a given flow, influent concentration, and polymer, the capital and operating costs depend almost entirely on these two variables. The operating costs are determined primarily by the polymer exhaustion rate because this will set the frequency of regeneration required. Total capital costs will be determined primarily by the superficial liquid retention time because this will set the size (or number) of the resin beds required.

For the purpose of cost estimating, data are generated for a phenol removal and recovery system. A flow diagram of this system and its operating characteristics follow.


SOLVENT MAKEUP
③
RECOVERED SOLVENT
PHENOLIC WASTE
①
②
WATER
PHENOL/WATER AZEOTROPE (RECYCLE TO ADSORBENT COLUMN)
POLYMERIC ADSORBER 1 (LOADING)
POLYMERIC ADSORBER 2 (REGENERATING)
DISTILLATION COLUMN #1 SOLVENT/WATER
DISTILLATION COLUMN #3 PHENOL/WATER
⑤
TREATED WASTE
⑥
WATER TO SEWER
RECOVERED PHENOL
④
DISTILLATION COLUMN #2 SOLVENT/WATER/PHENOL
SEPARATOR
70% PHENOL

COLUMN #3 UTILIZED WHEN HIGHLY PURE PHENOL IS REQUIRED

MATERIAL BALANCE FOR PHENOL REMOVAL AND RECOVERY SYSTEM
(lb/hr)

Material	Stream in flow diagram				
	1	2	3	4	5 plus 6
Phenol	264			264	<10 ppm
Water	21,736	1,480		9	23,207
Acetone			4		4
Total	22,000	1,480	4	273	23,211

The design basis for the recovery system included the following: (1) 1.2% phenol in water; (2) 330 d/yr operation, and (3) recovered pehnol of 2.08 x 10^6 lb/yr.

Fixed capital costs based on an ENR index of 3119 are estimated to be $1,060,000. Similarly, total capital cost is estimated to be $1,120,000. Estimation of annual operating cost is presented in the following table.

Cost item	Annual quantity	Cost per unit quantity	Annual cost,[a] $	
Direct operating cost				
Labor				
Operating	4,500 man-hr	$16/hr	72,000	
Maintenance			21,000	
				93,000
Power	150 x 10^3 kWh	$0.03/kWh		4,500
Steam	42 x 10^6 lb	$5/MM Btu		210,000
Water				
Makeup water	125 x 10^6 gal	$0.57/1,000 gal	71,000	
Cooling water	250 x 10^6 gal	$0.04/1,000 gal	11,000	
				82,000
Materials				21,500
Chemicals				
Resin replacement	100 ft^3	$175/$ft^3$	17,500	
Acetone replacement	35,000 lb	$0.24/lb	8,400	
				25,900
Total				437,000
Total indirect operating cost				355,000
Total annual operating cost				792,000

[a] Excludes annual credit based on selling phenol.

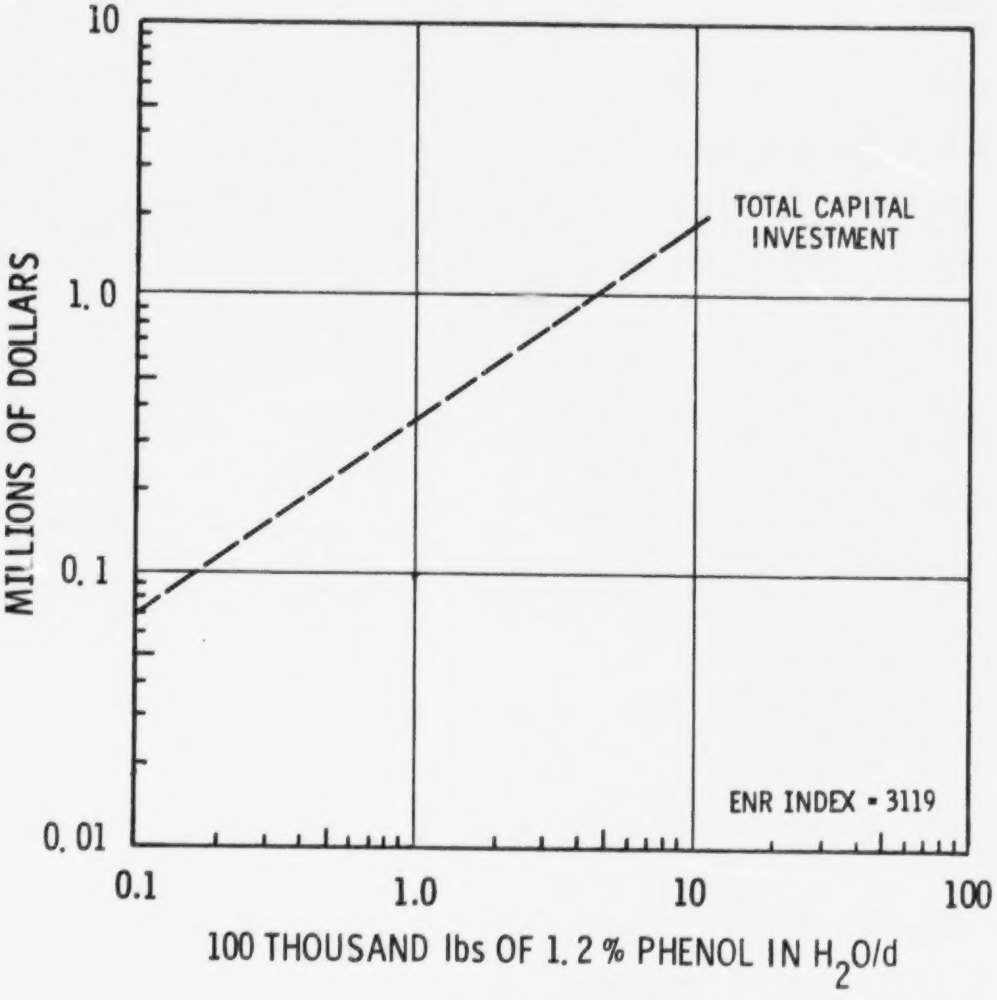
Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying flowrates.

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power, chemicals, materials, steam and water	1.0

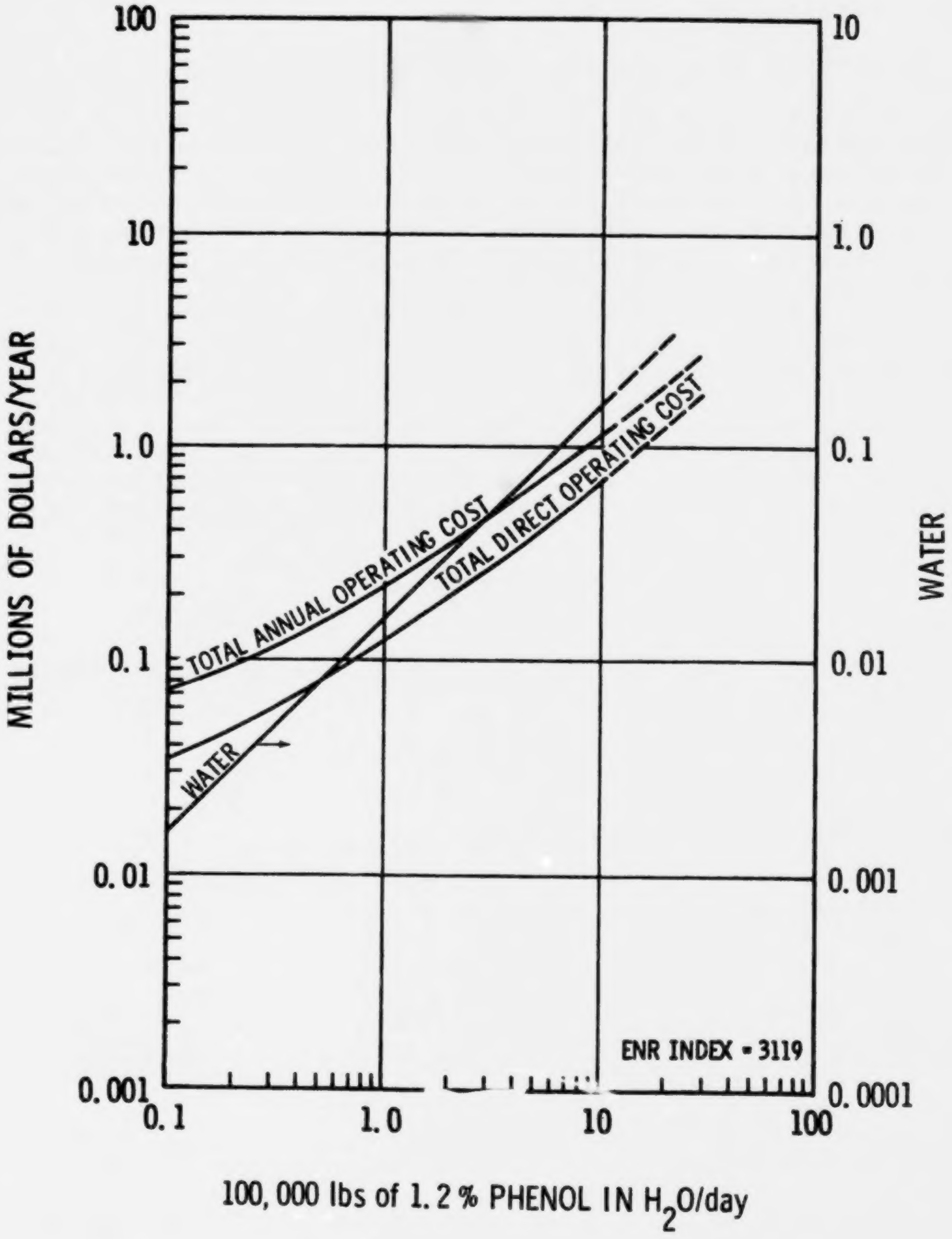
IV.5.8.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 2-1 to 2-26.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



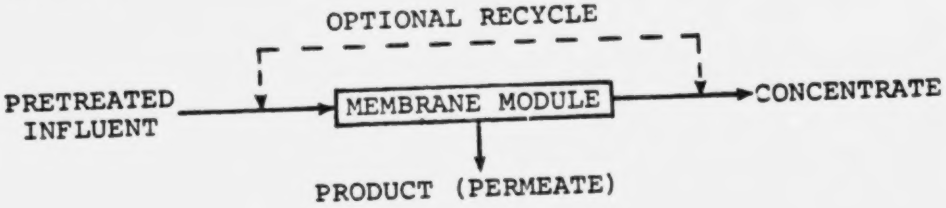
IV.5.9. REVERSE OSMOSIS [1]

IV.5.9.1 Description

Reverse osmosis is used for the removal of dissolved organic and inorganic materials and control of such wastewater parameters as soluble metals, TDS, and TOC.

Reverse osmosis (RO) separates dissolved materials in solution by filtration through a semipermeable membrane at a pressure greater than the osmotic pressure caused by the dissolved materials in the wastewater. With existing membranes and equipment, operating pressures vary from atmospheric to 1,500 psi. Products from the process are (1) the permeate or product stream with dissolved material removed, and (2) concentrate stream containing all removed material. Removal levels obtainable are dependent on membrane type, operating pressure, and the specific pollutant of concern. Removal of multicharged cations and anions is normally very high, while most low molecular weight dissolved organics are not removed or are only partially removed.

Typical equipment includes: membrane modules; feed, product, concentrate tanks; high pressure pump; prefilter plus pump; stainless steel piping; heat exchanger; flow and pressure instrumentation. An overall flow diagram of an RO system is shown below.



IV.5.9.2 Typical Design Criteria

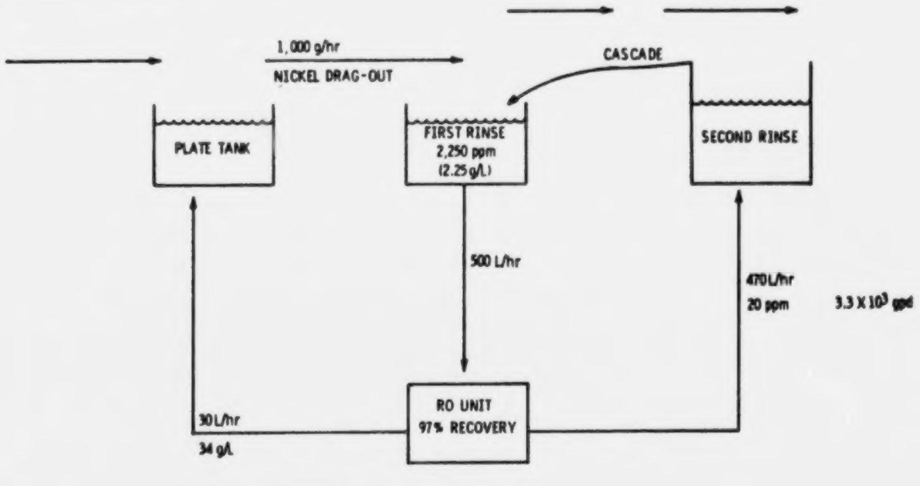
Typical design criteria for the reverse osmosis process are summarized on the next page.

Design criteria	Value
Membrane type	Cellulose acetate (also di- and tri-acetate), polyamide, and polysulfone
Flux (product) rate at 600 psi, 5,000 ppm NaCl solution, and 25°C	6 to 10 gpd/ft^2 membrane or 25 to 100 gpd/ft^3 module
Rejection at 600 psi, 5,000 ppm NaCl solution, and 25°C	70% to 99% depending on membrane specification
Operating pressure	250 to 1,500 psi
Membrane configuration	Plate, tubular, spiral or hollow fiber
Water recovery	50% to 85% depending on minimum solubility

IV.5.9.3 Costs

Costs of reverse osmosis plants are strong functions of the purified-water output capacity. They vary considerably with size of plant and model chosen. Hence, the reader is cautioned against generalization of the associated cost curves.

For the purpose of cost estimating, data are generated for an RO unit which is used on cascaded rinses of an acid nickel plating line. A flow diagram of the unit and its operating characteristics are shown below.


1,000 g/hr
NICKEL DRAG-OUT
CASCADE
PLATE TANK
FIRST RINSE
2,250 ppm
(2.25 g/L)
SECOND RINSE
500 L/hr
470 L/hr
20 ppm
3.3 X 10^3 gpd
30 L/hr
34 g/L
RO UNIT
97% RECOVERY

* ASSUME LIQUID VOLUME OF DRAG-IN EQUALS DRAG-OUT.

The design basis for reverse osmosis included 24-hr/d, 330-d/yr operation; pumps and filter, and cellulose acetate membranes in a spiral configuration.

Fixed capital costs based on an ENR index of 3119 are estimated to be $16,000. Similarly, total capital cost is estimated to be $17,000. Estimation of annual operating cost is presented in the following table based on a flowrate of 3,300 gal/d.

Cost item	Annual cost, $	
Direct operating cost		
Labor	1,320	(1/4-hr/d @$16/hr)
Membrane replacement	436	(replaced @2.5 yr)
Power	375	($0.035/kWh)
Total	2,131	
Annual operating cost	5,491	

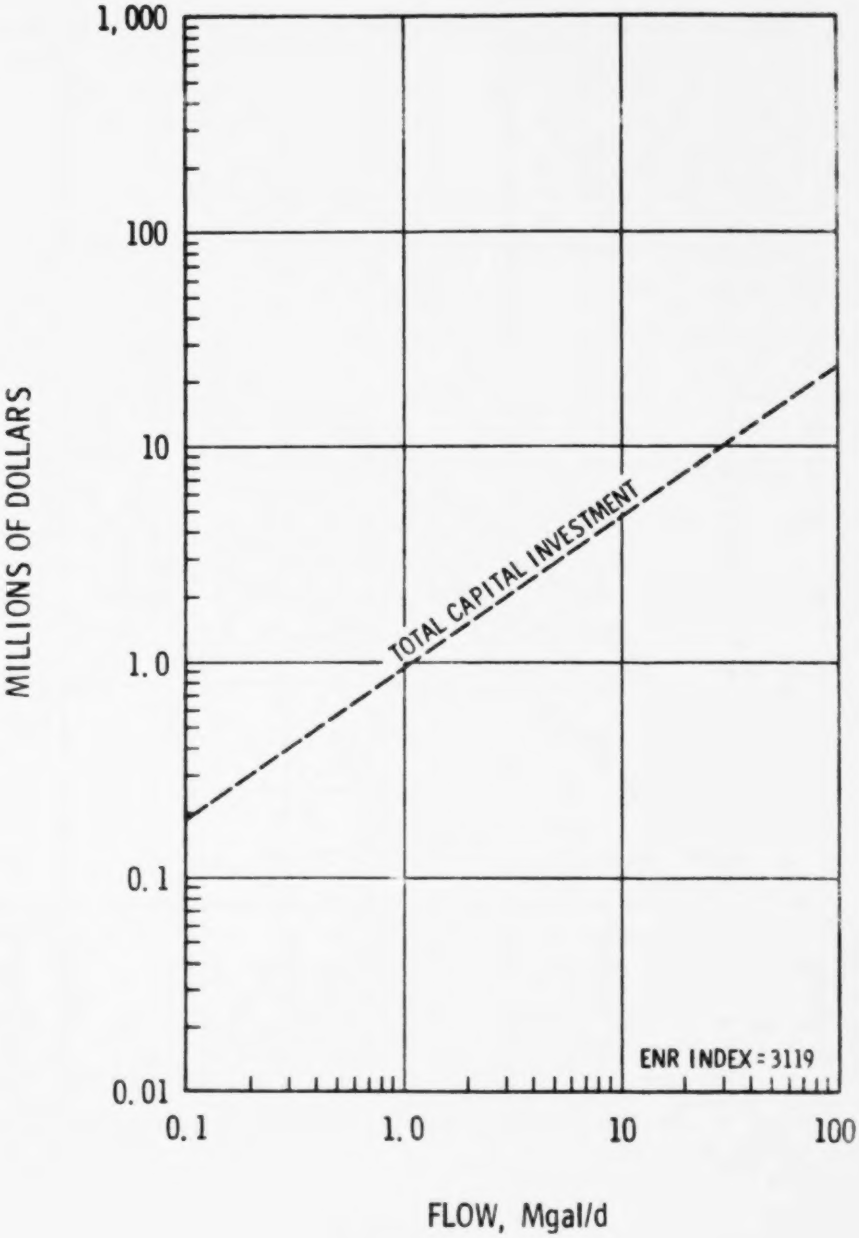
Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying flowrates:

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power and chemicals	1.0

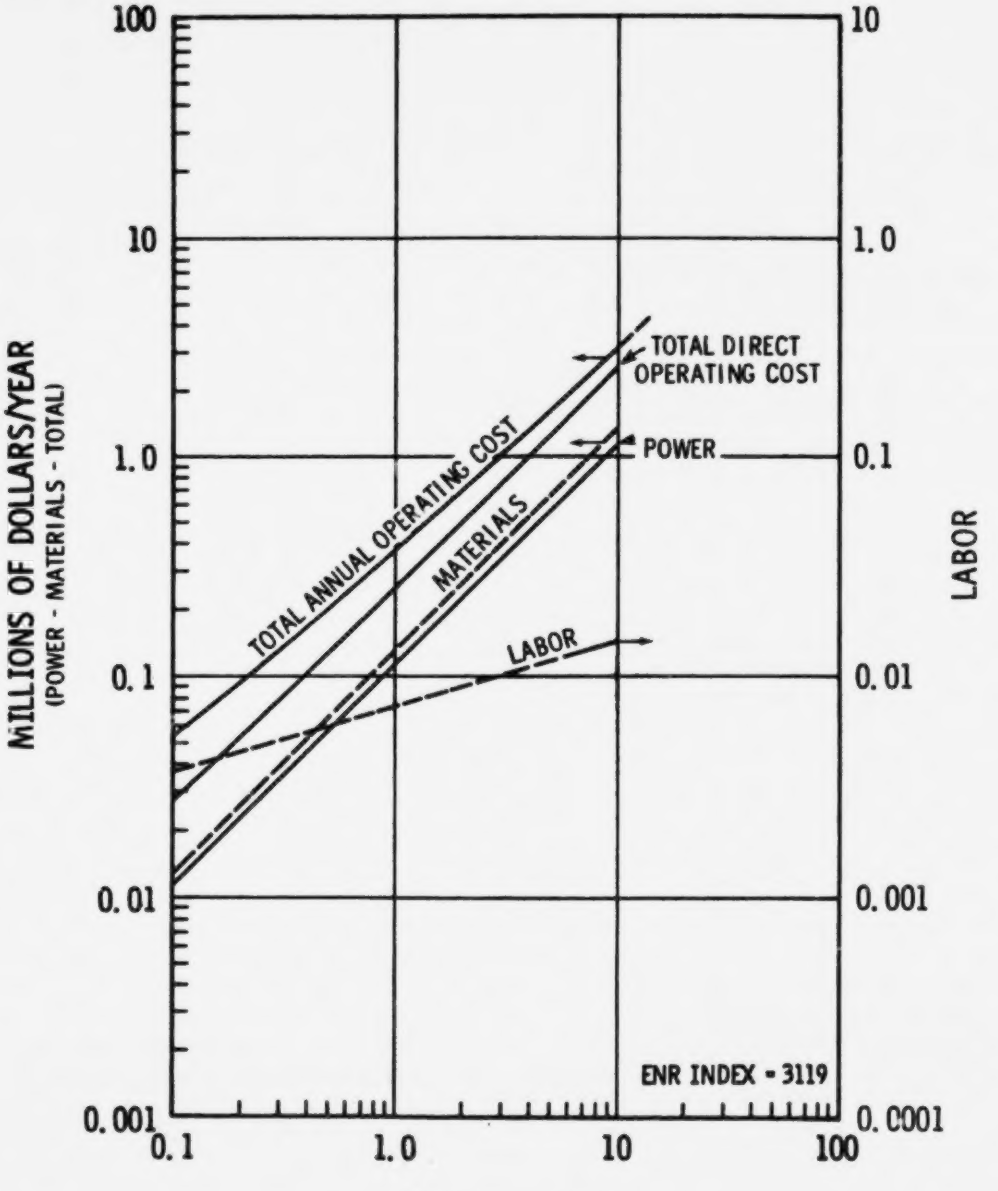
IV.5.9.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 39-1 to 39-19.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.10 ELECTRODIALYSIS [1]

IV.5.10.1 Description

The general function of electrodialysis is the separation of an aqueous stream under the action of an electric field into two streams: an enriched stream (more concentrated in electrolyte than the original), and a depleted stream. Success of the process depends on special synthetic membranes, usually based on ion exchange resins, which are permeable only to a single charge type of ion. Cation exchange membranes permit passage only of positive ions under the influence of the electric field; anion exchange membranes permit passage only of negatively charged ions.

In the electrodialysis process, feed water passes through compartments formed by the spaces between alternating cation-permeable and anion-permeable membranes held in a stack. At each end of the stack is an electrode that has the same area as the membranes. A dc potential applied across the stack causes the positive and negative ions to migrate in opposite directions. Because of the semipermeability of the membranes, a given ion will either migrate to the adjacent compartment or be confined to its original compartment, depending on whether or not the first membrane it encounters is permeable to it. As a result, salts are concentrated or diluted in alternate compartments.

To achieve high throughput, electrodialysis cells in practice are made very thin and assembled in stacks of cells in series. Each stack often consists of more than 100 cells. Feed material is first filtered to remove suspended particulate matter that could clog the system or foul the membrane and, if required, is given a pretreatment to remove oxidizing materials and ferrous or manganous ions, which would damage the membranes. Very high organic levels may also lead to membrane fouling. The catholyte stream is commonly acidified to offset the increase in pH that would normally occur within the cell, and an antiscaling additive may be required as well. An operating plant usually contains many recirculation, feedback, and control loops and pumps to optimize the concentrations and pH's at different points and thus maximize the overall efficiency. Although a certain amount of water transfer (electroosmosis) does occur, the process can be categorized with ion exchange, solvent extraction, or adsorbent processes as one in which solutes are removed from the solvent, rather than with distillation, freezing, or reverse osmosis in which the solvent is transported.

All ionized species are not removed in proportion to their concentration because of different mobilities and equilibrium concentrations within the membrane. Therefore, a solution partially deionized or concentrated by electrodialysis may contain significantly different proportions of ionized species than does the original feed.

Many colloids and polyanions have a net negative charge. For this reason they may collect upon or foul anion exchange membranes because of their positively charged functicnal groups. This problem may be avoided to some extent using an electrodialysis cell that consists of alternating cation and "neutral" membranes. Such systems utilizing a porous "neutral" membrane to avoid convective flow or mixing, frequently perform very well from a separation standpoint although they are not common commercially because of their higher electrical power requirements.

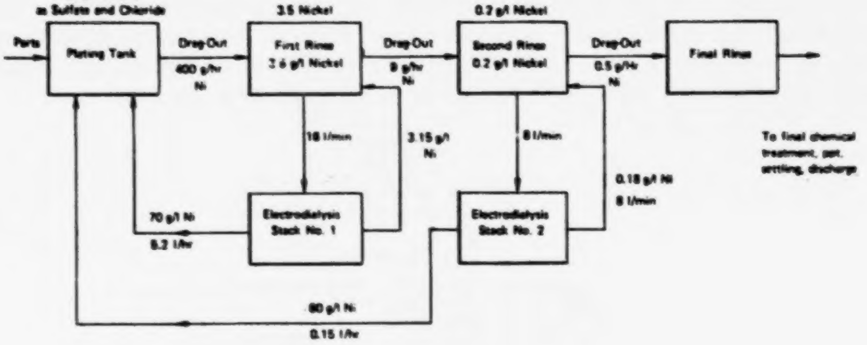
Generally, electrodialysis works best on acidic streams containing a single principal metal ion (such as acid nickel baths). At alkaline pH's, membrane life may diminish, but the system has been reported useable up to pH 14 under special circumstances. Mixed metals may not be concentrated in the same ratio as that in the feed, leading to problems in recycle. In addition, although a sodium and copper cyanide stream may perform as expected under electrodialysis, the presence of zinc (a common occurrence, especially in brass plating) can foul the anion membrane by the $(ZnCl)^-$ ion and partially convert that membrane to the cation form, with significant loss in system performance. If strongly alkaline, the feed streams are generally neutralized or rendered slightly acidic to prevent degradation of the anion membrane, which usually contains quaternary ammonium groups. Iron and manganese in the feed water also degrade most common membranes and must be removed if their total concentration in the feed water is greater than about 0.3 mg/L.

Calcium sulfate scale can also accumulate if the calcium concentration in the concentrated stream is allowed to exceed about 400 mg/L. Addition of a sequestering agent to the feed permits operation to a higher calcium concentration, but generally not above 900 mg/L. For this reason, the brine rarely constitutes less than 10 to 15% of the feed water volume (a concentration factor of 6 to 10).

Because the process depends on electrolytic conductance through the various liquid streams, it is rarely practical to produce product water of less than about 250 ppm total dissolved solids. For the same reason, it is often desirable to operate an electrodialysis system at a slightly elevated temperature. As a rule of thumb, a temperature increase of 17°C reduces the power consumption by 1%.

Membrane life, although dependent upon service conditions, is frequently five years. Other conponents are generally long lived, because the system, although somewhat corrosive perhaps, operates at a modest or ambient temperatures and pressures, and abrasives and particulates normally will have been removed from the feed water.

A flow diagram of one electrodialysis system is shown below.

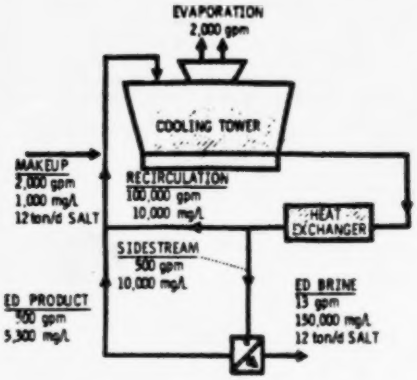

90 g/l Nickel,
as Sulfate and Chloride
Parts
Plating Tank
Drag-Out
400 g/hr
Ni
3.5 Nickel
First Rinse
3.6 g/l Nickel
Drag-Out
9 g/hr
Ni
0.2 g/l Nickel
Second Rinse
0.2 g/l Nickel
Drag-Out
0.5 g/hr
Ni
Final Rinse
18 l/min
3.15 g/l
Ni
8 l/min
To final chemical
treatment, ppt,
settling, discharge
0.18 g/l Ni
8 l/min
70 g/l Ni
5.2 l/hr
Electrodialysis
Stack No. 1
Electrodialysis
Stack No. 2
80 g/l Ni
0.15 l/hr

IV.5.10.2 Typical Design Criteria

Design criterion	Value
Flow rate	Max 10 MGD
Brine stream	10%-25% of feed stream
Membrane thickness	0.005 to 0.025 in
Plastic spacers	0.04 in thick
Temperature	≤100°F

IV.5.10.3 Costs

The costs for electrodialysis systems are dependent not only on the volume of water to be treated but also the amount of salts to be removed. The contributions of various economic factors is illustrated in the cost estimation for a cooling tower circuit with side stream electrodialysis. A flow diagram of this system and its operating characteristics are shown below.


EVAPORATION
2,000 gpm
COOLING TOWER
MAKEUP
2,000 gpm
1,000 mg/l
12 ton/d SALT
RECIRCULATION
100,000 gpm
10,000 mg/l
HEAT EXCHANGER
SIDESTREAM
500 gpm
10,000 mg/l
ED PRODUCT
500 gpm
5,300 mg/l
ED BRINE
13 gpm
150,000 mg/l
12 ton/d SALT

The design basis for the electrodialysis system includes the side stream electrodialysis of cooling tower water in a single stage with 300 d/yr operation.

Fixed capital costs based on an ENR index of 3119 are estimated to be $610,000. Similarly, total capital cost is estimated to be $646,000. Estimation of annual operating cost is presented in the following table.

Cost item	Annual quantity	Cost per unit quantity	Annual cost, $	
Direct operating cost				
Labor				
Operating	9,240 man-hr	$16/hr	150,000	
Maintenance			12,000	
				162,000
Chemicals				10,000
Materials				
Membrane replacement			9,700	
Maintenance			12,500	
				22,200
Power	6,000,000 kWh	$0.035/kWh		210,000
Total				404,200
Total indirect operating cost				240,000
Total annual operating cost				644,000

Cost curves were developed for total capital investment and annual operating cost using the operating and the cost data shown on the previous page and the following exponential scaling factors, which are used to determine costs at varying flowrates:

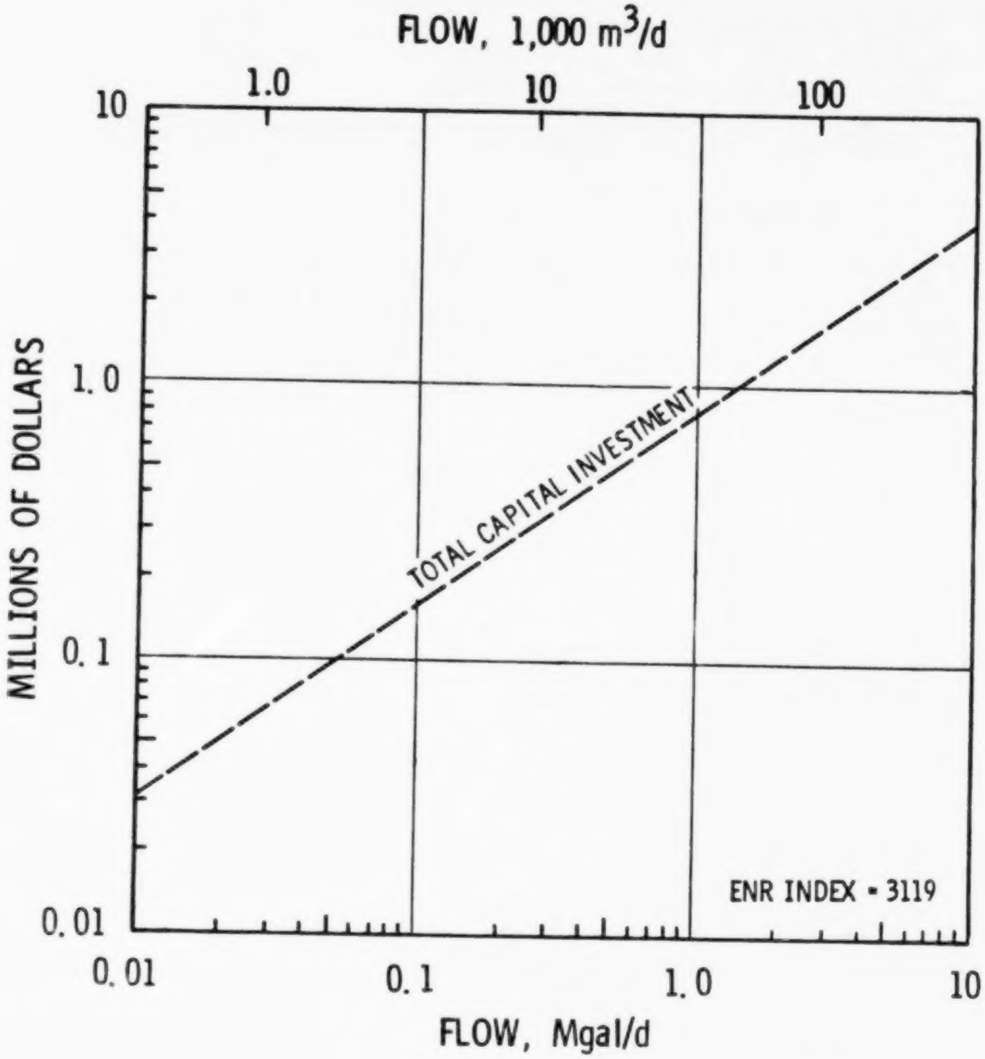
Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power, materials and chemicals	1.0

IV.5.10.4 References

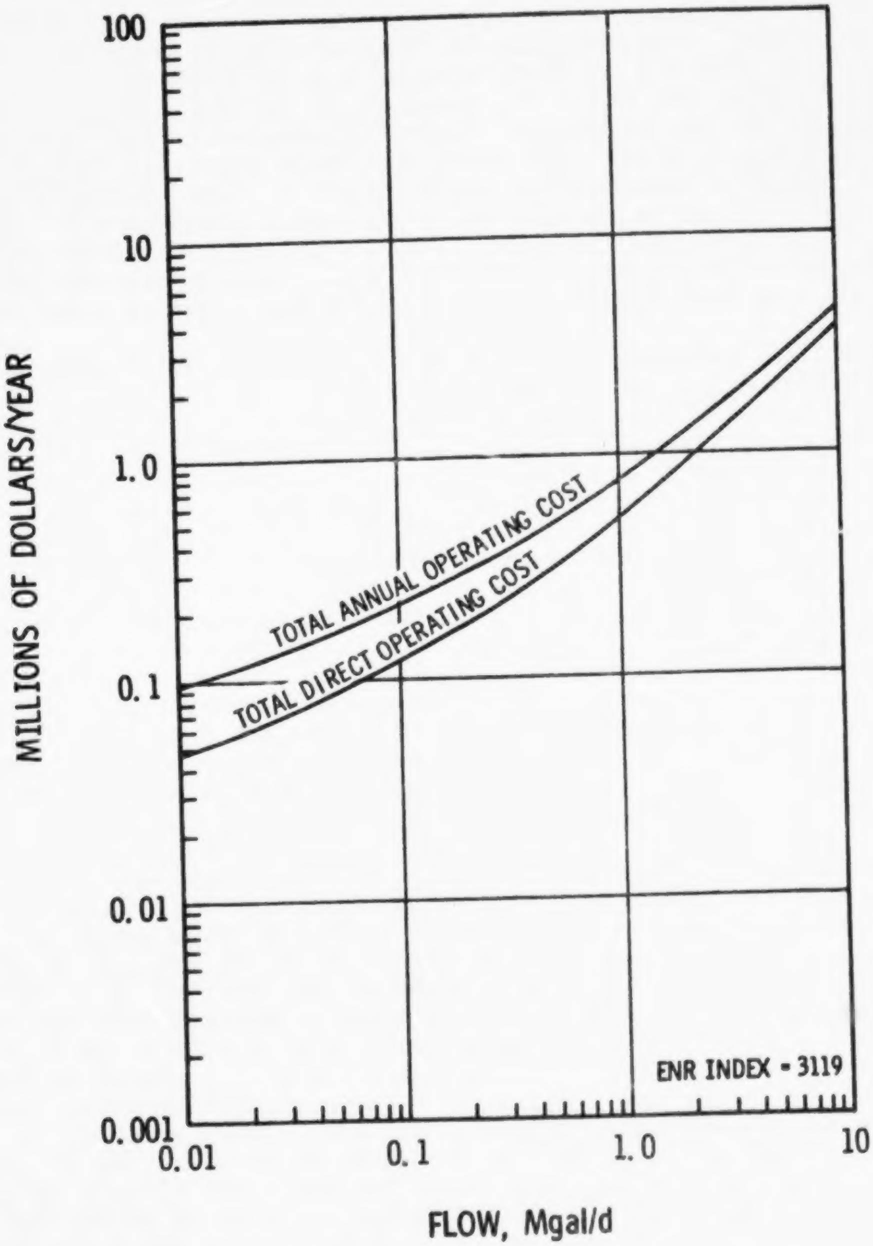
1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 18-1 through 18-14.

2. R.L. Culp; Wesner, G.M.; Culp, G.L. Handbook of Advanced Wastewater Treatment. Van Nostrand Reinhold Co. 1978 by Litton Educational Publishing Inc. New York, New York, 632 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.11 DISTILLATION [1]

IV.5.11.1 Description

Distillation is the boiling of a liquid solution and condensation of the vapor for the purpose of separating the components. In the distillation process, there are two phases: the liquid phase and the vapor phase. The components that are to be separated by distillation are present in both phases but in different concentrations. If there are only two components in the liquid, one concentrates in the condensed vapor (condensate) and the other in the residual liquid. If there are more than two components, the less volatile components concentrate in the residual liquid and the more volatile in the vapor or vapor condensate. The ease with which a component is vaporized is called its volatility, and the relative volatilities (ratio of equilibrium ratios) of the components determine their vapor-liquid equilibrium relationships.

There are five general types of distillation, and a general description of each type is provided below.

- Batch Distillation. The simplest form of distillation is a single equilibrium stage operation. It is carried out in a "still" in which the reboiler equivalent consists of a steam jacket or a heating coil. The liquid is "boiled"; the vapor is driven off, condensed, and collected in an accumulator (a condensed vapor collector) until the desired concentration of the "product" has been reached. As the remaining liquid becomes leaner in the volatile component and richer in the less volatile component, its volume diminishes. If the residual liquid is the product, then "bottoms" concentration will be the controlling parameter. The batch still, as previously described, consists of a vessel that provides one equilibrium stage. By adding a condenser and recycling some of the condensed vapor, a second vapor-liquid equilibrium stage is added, and the separation is improved.

- Continuous Fractional Distillation. In continuous fractional distillation, a steady stream feed enters the column, which contains plates or packing (packing is normally used only in small-scale equipment) that provide additional vapor-liquid contact (equilibrium) stages. Overhead vapors and bottoms are continuously withdrawn. Vapor from the top plate is condensed and collected in a vessel known as an accumulator. Some of the liquid in the accumulator is continuously returned to the top plate of the column as reflux while the remainder of the liquid is continuously withdrawn as the overhead product stream. At the bottom of the column, the liquid collects in the reboiler, where it is heated by steam coils or a steam jacket. The function of the reboiler is to receive the liquid overflow from the lowest plate and return a portion of this as a vapor stream, while the remainder is withdrawn continuously as a liquid bottom product.

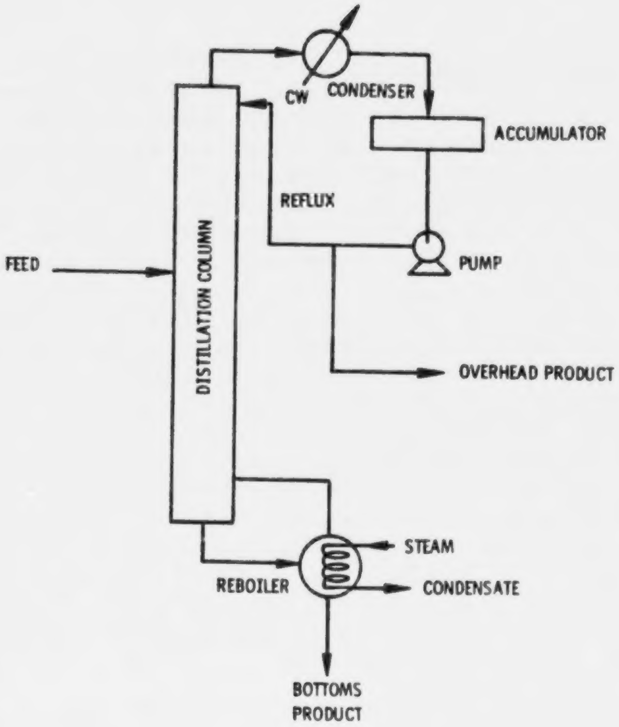
- Azeotropic Distillation. An azeotrope is a liquid mixture that maintains a constant boiling point and produces a vapor of the same composition as the mixture when boiled. Because the composition of the vapor produced from an azeotrope is the same as that of the liquid, an azeotrope may be boiled away at a constant pressure, without change in concentration in either liquid or vapor. Since the temperature cannot vary under these conditions, azeotropes are also called constant boiling mixtures.

An azeotrope cannot be separated by constant pressure distillation into its components. Furthermore, a mixture on one side of the azeotrope composition cannot be transformed by distillation to a mixture on the other side of the azeotrope. If the total pressure is changed, the azeotropic composition is usually shifted. Sometimes this principle can be applied to obtain separations under pressure or vacuum that cannot be obtained under atmospheric pressure conditions. Most often, however, a third component - an additive, sometimes called an entrainer - is added to the binary (two-component) mixture to form a new boiling-point azeotrope with one of the original constituents. The volatility of the new azeotrope is such that it may be easily separated from the other original constituents.

- Extractive Distillation. Extractive distillation is a multicomponent rectification method of distillation. A solvent is added to a binary mixture that is difficult or impossible to separate by ordinary means. This solvent alters the relative volatility of the original constituents, thus permitting separation. The added solvent is of low volatility and is not appreciably vaporized in the fractionator.

- Molecular Distillation. Molecular distillation is a form of a very low pressure distillation conducted at absolute pressures in the order of 0.003 mm of mercury that is suitable for heat-sensitive substances. Ordinarily, the net rate of evaporation is very low owing to the fact the evaporated molecules are reflected back to the liquid after collisions occurring in the vapor. By reducing the absolute pressure to values used in the molecular distillation, the mean free path of the molecules becomes very large (in the order of 1 cm). If the condensing surface is then placed at a distance not exceeding a few centimeters from the vaporing liquid surface, very few molecules will return to the liquid, and the net rate of evaporation is substantially improved.

A flow diagram of a distillation process is shown on the next page.


CW
CONDENSER
ACCUMULATOR
REFLUX
FEED
DISTILLATION COLUMN
PUMP
OVERHEAD PRODUCT
STEAM
REBOILER
CONDENSATE
BOTTOMS PRODUCT

IV.5.11.2 Typical Design Criteria

Typical Design Criteria is discussed in the Cost Section.

IV.5.11.3 Costs

To illustrate (but not typify) the economics of distillation, an economic analysis was performed on an "all purpose" utility distillation system. This type of system is most likely to be installed when there is a high probability that many different types of organic liquid waste streams will be treated. Because of the high degree of flexibility, the utility column is not designed for optimum operation with any particular stream or flow rate.

For the purpose of cost estimating, data are generated for the distillation of 20% acetone in a water waste stream. A flow diagram of this distillation system showing the equipment and operating characteristics assumed is presented below.

Operating characteristic	Value
Column[a]	
Diameter	3 ft
Height	80 ft
Packing	Berl saddles; 1 1/2-in. stone
Liquid redistribution plates	Located every 4 ft
Reboiler	375 ft^2
Condenser	100 ft^2
Pumps	Centrifugal; 5 gpm ratio and 1/2 hp motor
Material of construction	Stainless steel

Conditions at equilibrium:

Stream[b]	Flow rate, lb/hr	Temperature, °C	Concentration, wt %	
			Acetone	Water
Feed (F)	1,000	21	20	90
Overheads (0)	812	59	98.5	1.5
Reflux (R)	609	59	98.5	1.5
Product heads (H)	203	59	98.5	1.5
Bottom (B)	797	100	0	100

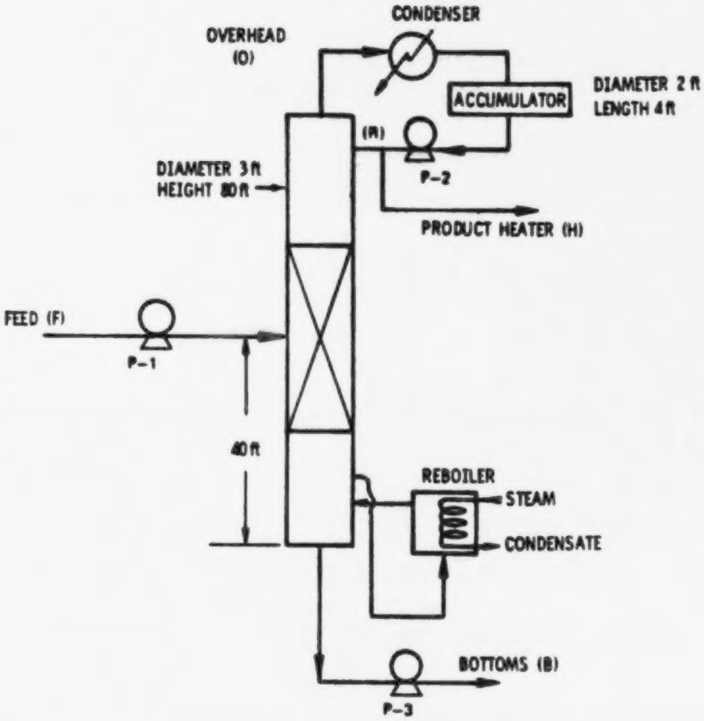
[a] Column operates under slight positive pressure; reflux ratio = 3.1.

[b] Letter in ( ) corresponds to those on flow diagram.

The design basis for distillation included 8,000 lb waste/d; 125 d/yr, 8 hr/d operation, and 50% of utility column utilization (actual system).

Fixed capital costs based on an ENR index of 3119 are estimated to be $312,000. Similarly, total capital cost is estimated to be $330,000. Estimation of annual operating cost is presented in the following table.

Cost item	Annual quantity	Cost/unit	Annual cost, $	
Direct operating cost				
Labor				
Operating	375 man-hr	16/hr	6,000	
Maintenance			2,300	
				8,300
Water				
Makeup	6,000 gal	$0.57/1,000 gal	3	
Cooling	535,000 gal	$0.04/1,000 gal	23	
				26
Steam	1,000 MM Btu	$5/MM Btu		5,000
Materials				2,400
Power	290 kWh	$0.035/kWh		9
Total				16,000
Total indirect operating cost				47,900
Total annual operating cost				64,000


OVERHEAD
(O)
CONDENSER
ACCUMULATOR
DIAMETER 2 ft
LENGTH 4 ft
(R)
P-2
DIAMETER 3 ft
HEIGHT 80 ft
PRODUCT HEATER (H)
FEED (F)
P-1
40 ft
REBOILER
STEAM
CONDENSATE
BOTTOMS (B)
P-3

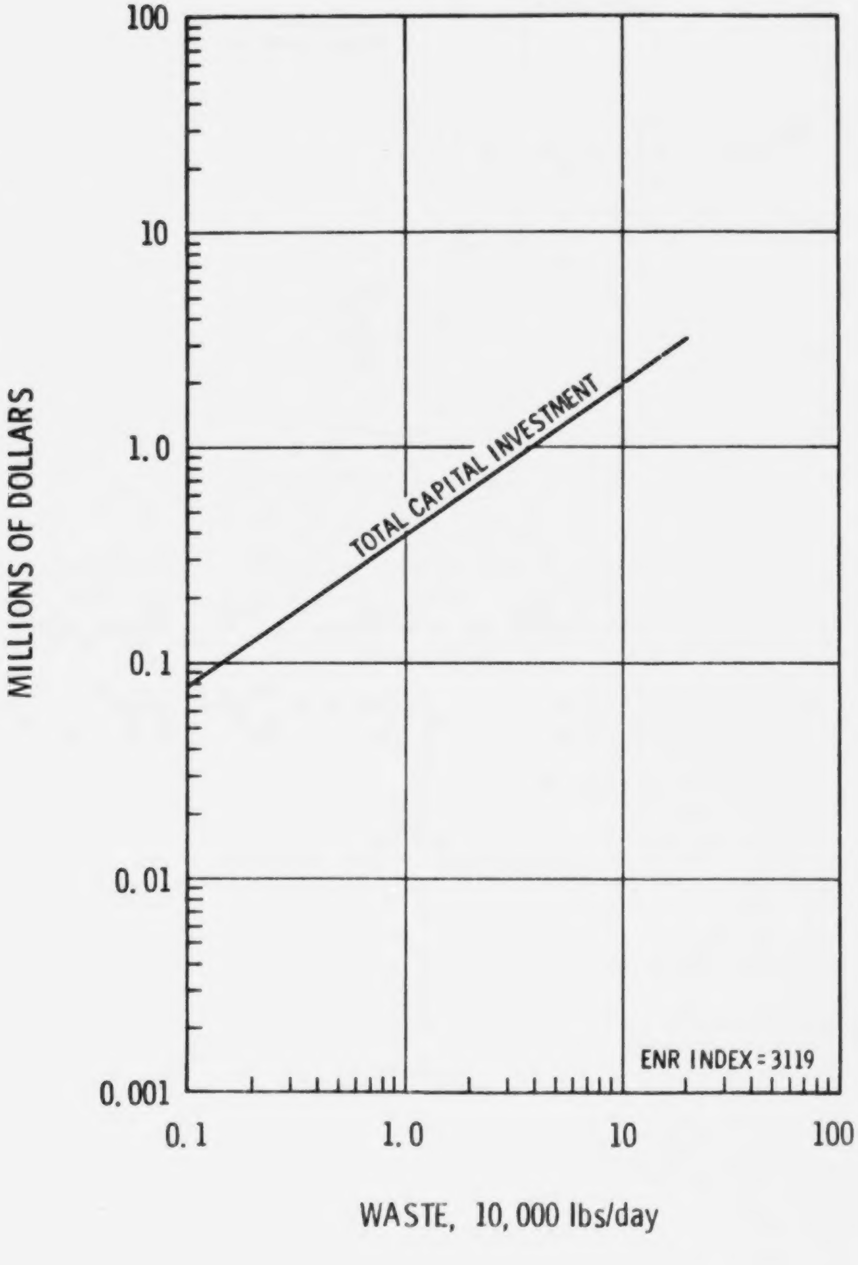
Cost curves were developed for total capital investment and annual operating cost using the cost data shown above and the following exponential scaling factors, which were used to determine costs at varying flowrates.

Cost item	Exponential factor
Total capital investment	1.2
Labor	0.3
Power, chemicals, water and steam	1.0

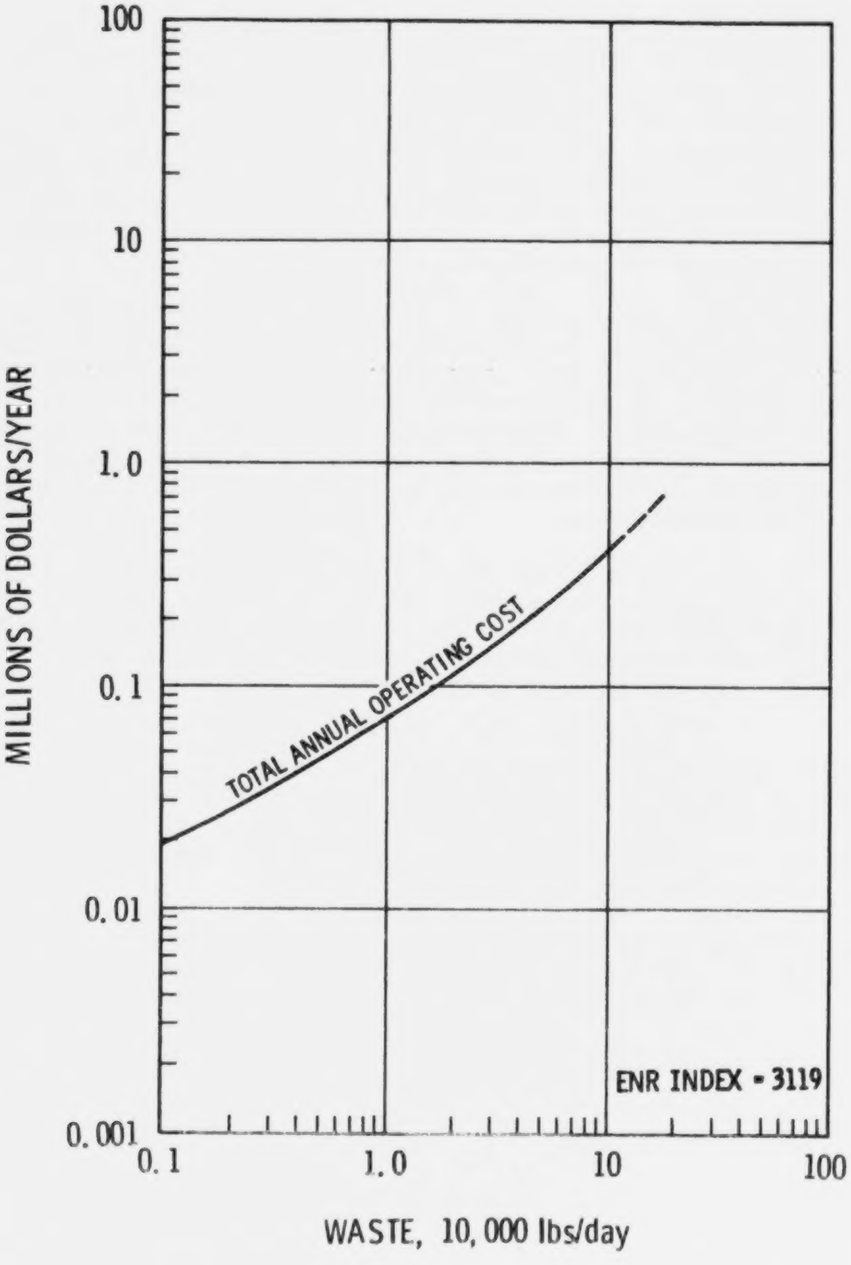
IV.5.11.4 References

1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 17-1 to 17-35.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.12 CHLORINATION (DISINFECTION) [1]

IV.5.12.1 Description

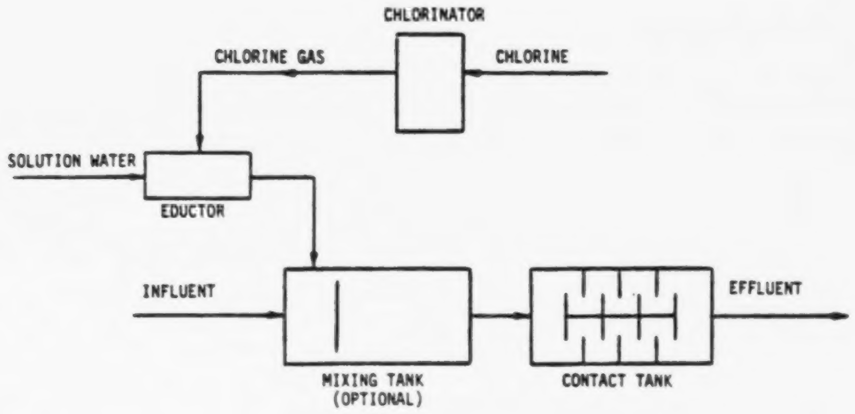
Chlorination is the most commonly used disinfection process; it is especially used for the removal of pathogens and other disease causing organisms. Chlorination involves the addition of elemental chlorine or hypochlorites to the wastewater. Chlorine combines with water to form hypochlorous (HOCl) and hydrochloric (HCl) acids. Hydrolysis goes virtually to completion at pH values and concentrations normally experienced in municipal wastewater applications. Hypochlorous acid will ionize to hypochlorite (OCl) ion significantly above a pH of 7.5. Both are considered free residual chlorine with hypochlorous acid having the greater disinfecting power in water. In wastewater, the primary disinfectant species is monochloromine.

The amount of chlorine added is determined by cylinder weight loss. Chlorine demand is determined by the difference between the chlorine added and the measured residual concentration after a certain period has passed from the time of addition; this is usually 15-30 minutes. The chlorine or hypochlorite is rapidly mixed with the wastewater, after which it passes through a detention tank, which normally contains baffled zones to prevent short circuiting of wastewater.

Chlorine is also used as an oxidizing agent for taste and odor control in wastewater treatment plants and for cyanide destruction in industrial applications. One of the major problems with the use of chlorine for disinfection, taste, and odor control is the formation of chlorinated organics, many of them priority pollutants.

Chlorine or hypochlorite salts can be used. The two most common hypochlorite salts are calcium and sodium hypochlorite. Dechlorination may be used; this generally involves the addition of sulfur dioxide, aeration, or even activated carbon, when chlorine residual standards are strict.

Equipment normally associated with the chlorination process includes chlorine analyzers, pH controllers, chemical feeders, and mixers. A flow diagram of a chlorination system is shown on the next page.



IV.5.12.2 Typical Design Criteria

Generally a contact period of 15 to 30 minutes at peak flow is required. Detention tanks should be designed to prevent short circuiting; this usually involves the use of baffling. Baffles can either be the over-and-under or the end-around varieties. Residuals of at least 0.5 mg/L are generally required. The following table presents typical dosages for disinfection:

Effluent source	Dosage range, mg/L
Untreated wastewater (prechlorination)	6 - 25
Primary sedimentation	5 - 20
Chemical-precipitation plant	3 - 10
Trickling-filter plant	3 - 10
Activated-sludge plant	2 - 8
Multimedia filter following activated-sludge plant	1 - 5

IV.5.12.3 Costs

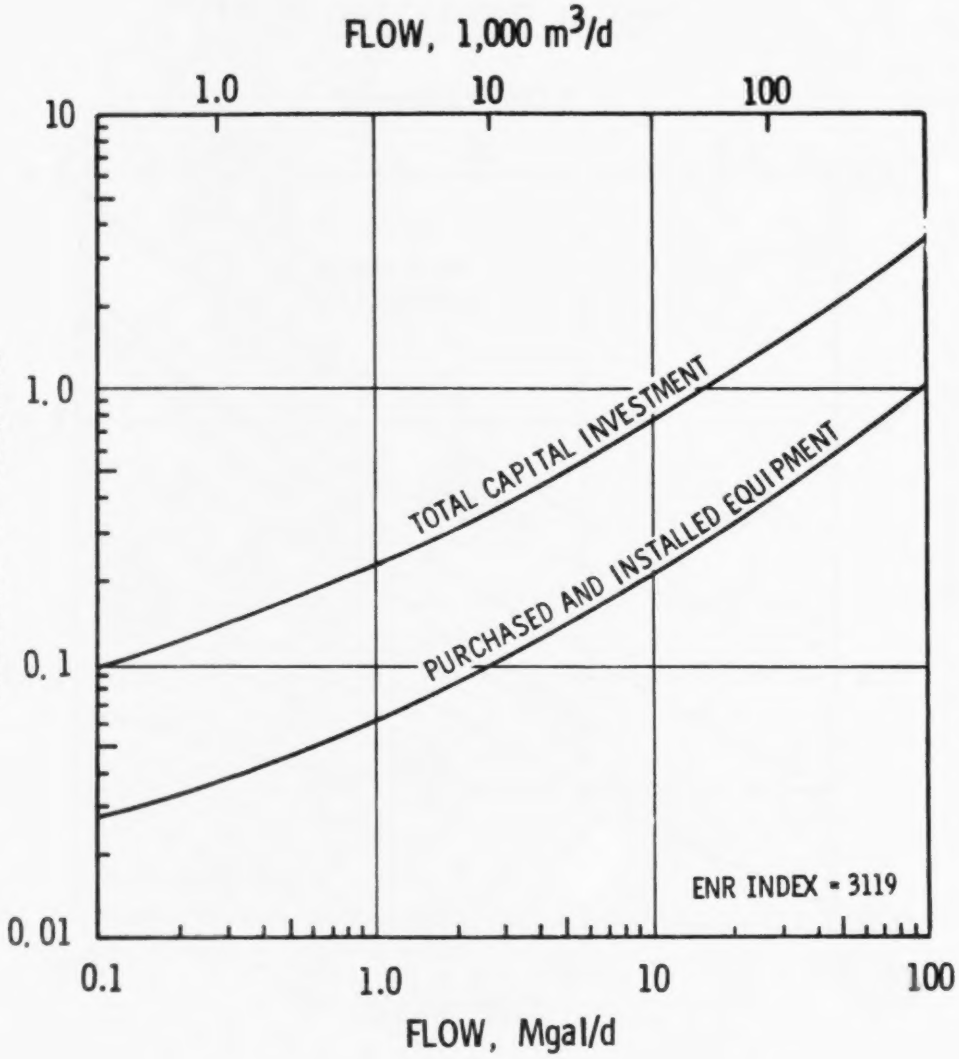
Purchased equipment and installation cost for estimation of the total capital investment includes chlorine supply, chlorinator, and contact chamber. Chlorine cost is assumed to be $160/ton. The operating characteristics shown below were assumed for cost estimation.

Operating characteristic	Assumed value
Service life	15 yr
Dosage	10 mg/L
Contact time	30 min

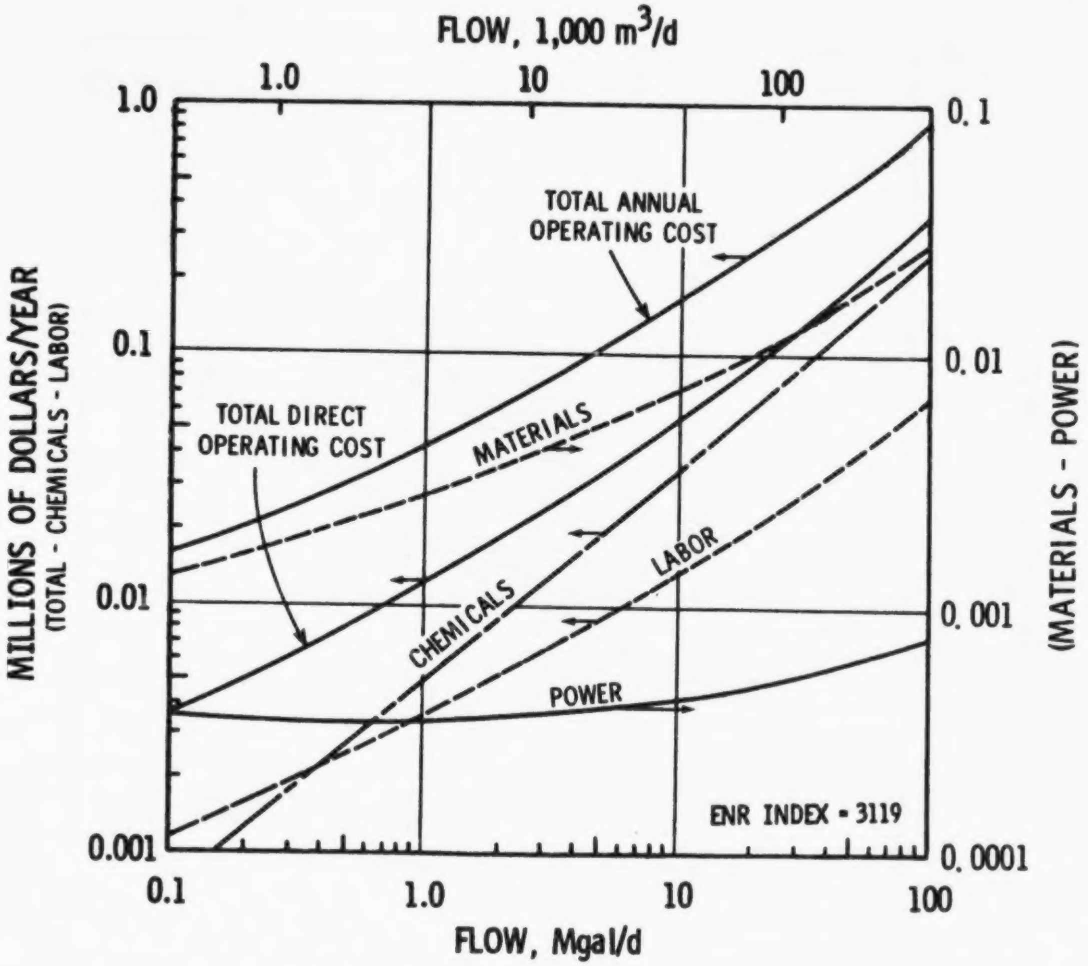
IV 5.12.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.13 DECHLORINATION [1]

IV.5.13.1 Description

Dechlorination is used to remove free and combined chlorine from effluents in order to reduce the toxic effects of chlorinated effluents. Both free chlorine and chloramine residuals are toxic to fish and other aquatic organisms.

Therefore, dechlorination involves the addition of sulfur dioxide to wastewater, whereby the following reactions occur:

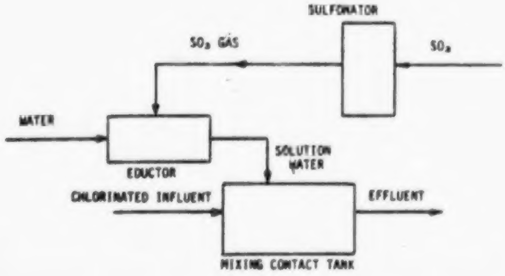
$$SO_2 + HOCl + H_2O = SO_4^{-2} + Cl^- + 3H^+ \text{ (For free chlorine)} \quad (1)$$

$$SO_2 + NH_2Cl + 2H_2O = SO_4^{-2} + Cl^- + 2H^+ + NH_4^+ \text{ (For combined chlorine)} \quad (2)$$

As noted, small amounts of sulfuric and hydrochloric acids are formed; however, they are generally neutralized by the buffering capacity of the wastewater. Dechlorination can also be used in conjunction with superchlorination. Because superchlorination involves the addition of excess chlorine, dechlorination is required to eliminate this residual. Sulfur dioxide, the most common chemical used for dechlorination, is fed as a gas, using the same equipment as chlorine systems. Because the reaction of sulfur dioxide with free or combined chlorine is practically instantaneous, the design of contact systems is less critical than that of chlorine contact systems. Detention of less than 5 minutes is quite adequate, and in-line feed arrangements may also be acceptable under certain conditions.

Metabisulfite, bisulfite, or sulfite salts can be used, as can automatic or manually fed systems. If chlorine is used at the site, sulfur dioxide is preferred, because identical equipment can be used for the addition of both chemicals. Alternative dechlorination systems include activated carbon, H_2O_2, and ponds (sunlight and aeration).

Equipment normally associated with the dechlorination process includes chemical feeders, mixers, and automatic controls. A flow diagram of a dechlorination system is shown below.



IV.5.13.2 Typical Design Criteria

Typical design criteria for a dechlorination system are shown below.

Design criterion	Value
Contact time	1 - 5 min
Sulfur dioxide feed rate	1.1 lb/lb residual chlorine
Sodium sulfite feed rate	0.57 lb/lb chlorine
Sodium bisulfite feed rate	0.68 lb/lb chlorine
Sodium thiosulfate feed rate	1.43 lb/lb chlorine

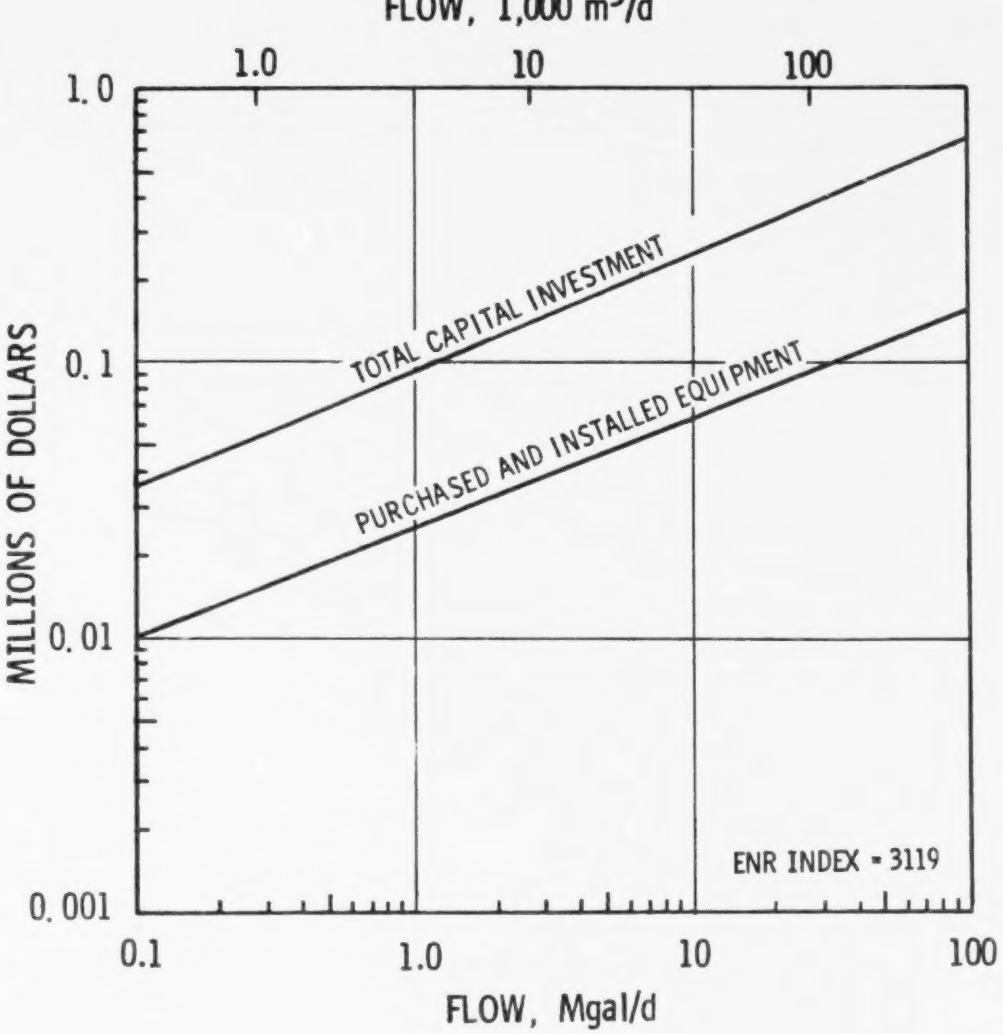
IV.5.13.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes SO_2 feed facilities, reaction tank (1-min detention time), mixer, and storage facilities. Building space and control instrumentation are not included. SO_2 costs were based on 20 lb/Mgal (1.1 mg/L of SO_2 required per mg/L of chlorine residual).

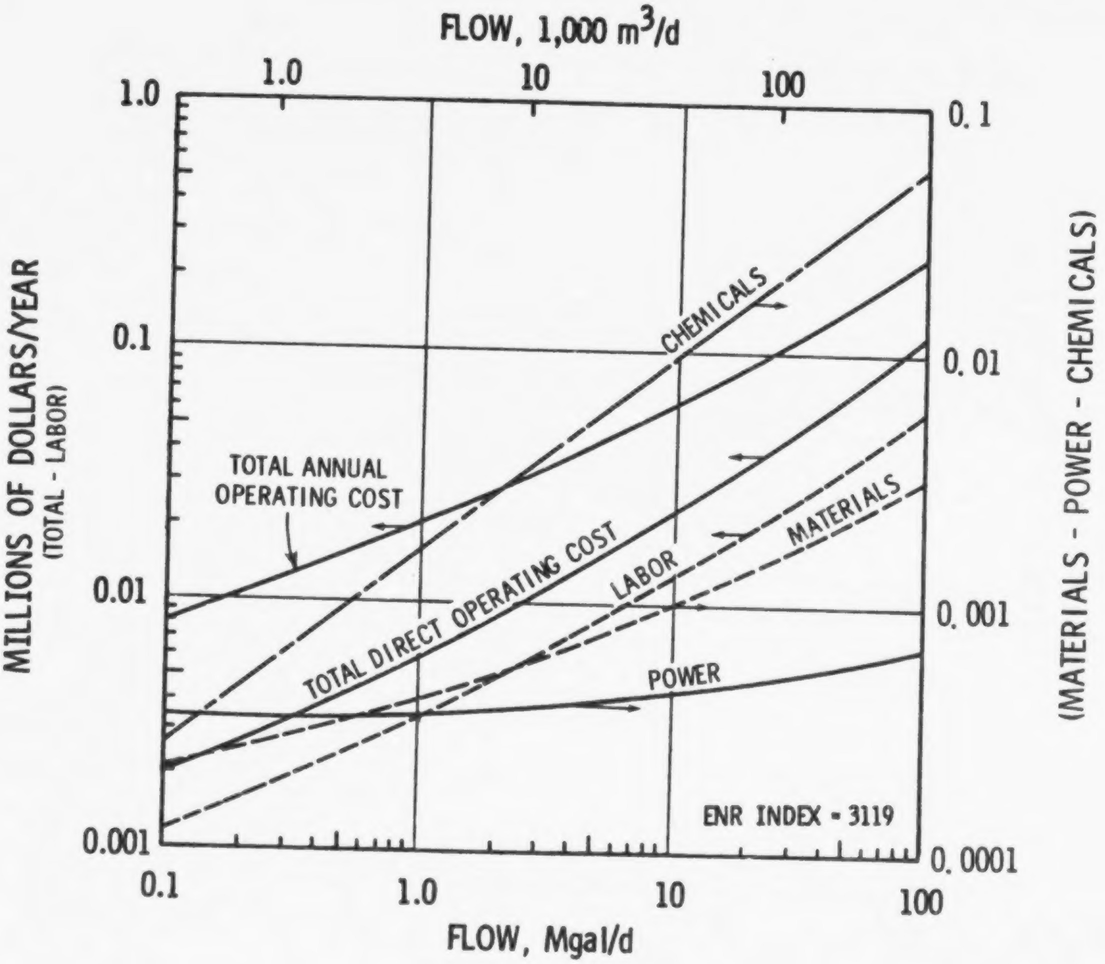
IV.5.13.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.5.14 OZONATION [1]

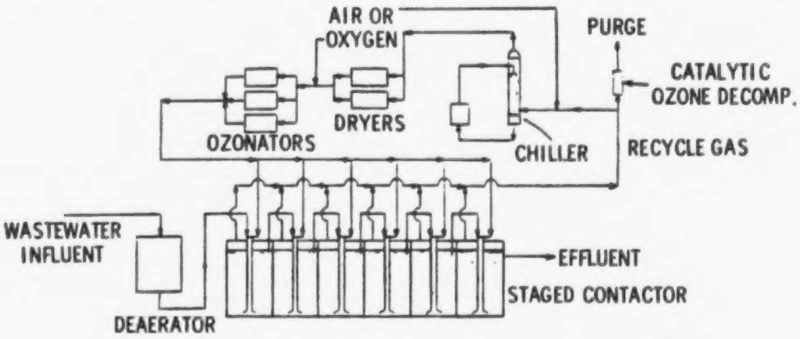
IV.5.14.1 Description

Ozone (O_3) is a very strong oxidant, which may be used, at dosages of 10 to 300 mg/L, to remove residual dissolved organics from secondary effluent. Ozone has also been experimentally used to treat raw wastewater and wastewater after various stages of treatment. The rate of oxidation is both temperature- and pH-dependent. Reaction rates increase with increasing temperature. The optimum pH range is 6 to 8. Because there is a wide range of ozone reactivity with the diverse organic content of wastewater, both the required ozone dose and reaction time are dependent on the quality of the influent to the ozonation process. Generally, higher doses and longer contact times are required for ozone oxidation reactions than are required for wastewater disinfection using ozone.

Systems have been designed to utilize staged contactors (injection type) with recycle of the ozone/oxygen off-gas. On these systems, provision must be made for the removal of nitrogen gas from the influent wastewater stream and for the removal of reaction-produced carbon dioxide from the off-gas stream in those situations that warrant it. Ozone tertiary treatment may eliminate the need for a final disinfection step.

Ozone breaks down to elemental oxygen in a relatively short period of time (half life about 20 minutes). Consequently, it must be generated on site using either air or oxygen as the feed gas. Ozone generation utilizes a silent electric arc or corona through which air or oxygen passes, and yields ozone in an air/oxygen mixture. The percentage of ozone formed is a function of voltage, frequency, gas flow rate, and moisture. Automatic devices are commonly applied to control and adjust the ozone generation rate.

Equipment normally associated with the ozone oxidation process includes oxygen generator, columns-towers, ozone auxiliary equipment, and ozone generator. A flow diagram of the ozone oxidation system is shown below.



IV.5.14.2 Typical Design Criteria

Typical design criteria for an ozone oxidation system are shown below.

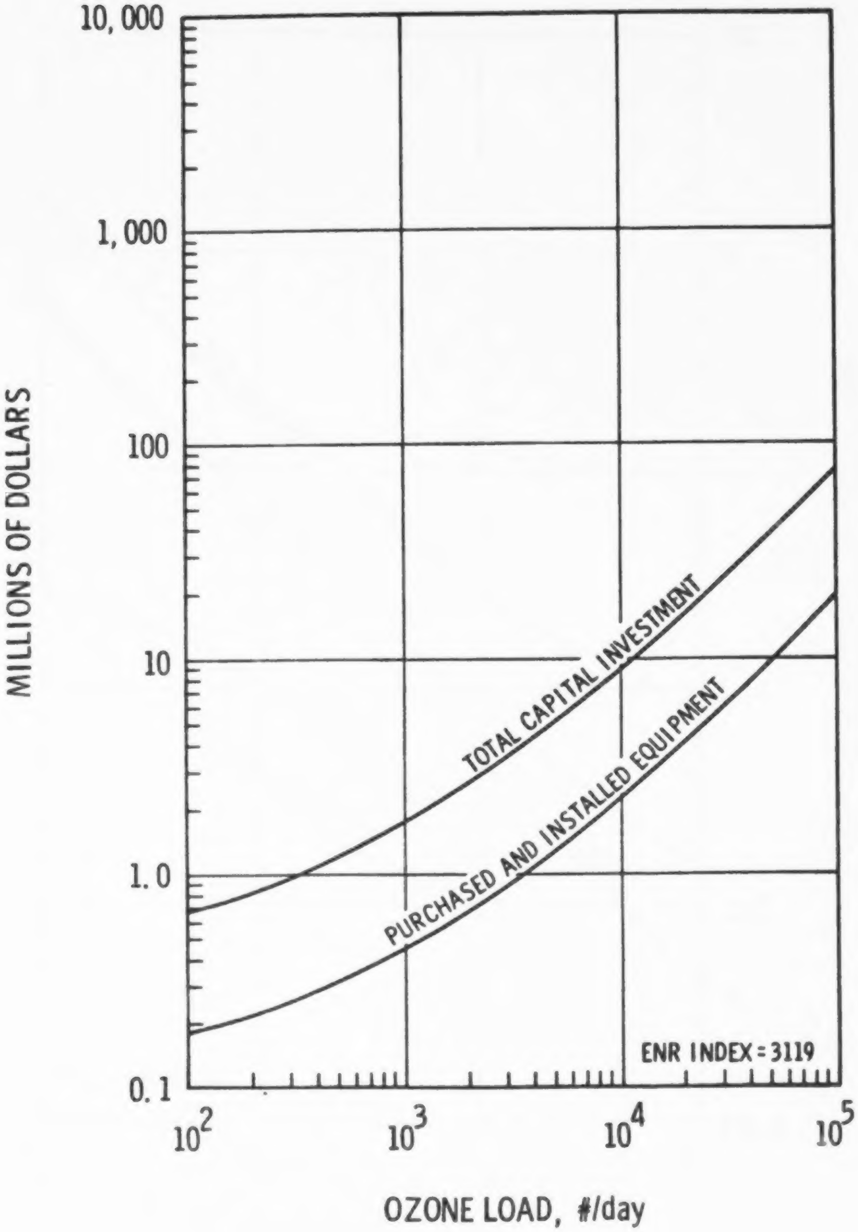
Design criterion	Value
Contact time	1 - 90 min
Dosage rate	10 - 300 mg/L
Ozone production	4.5 kWh/lb from oxygen 7.5 kWh/lb from air
pH range	5 - 11 (6 - 8 optimum)

IV.5.14.3 Costs

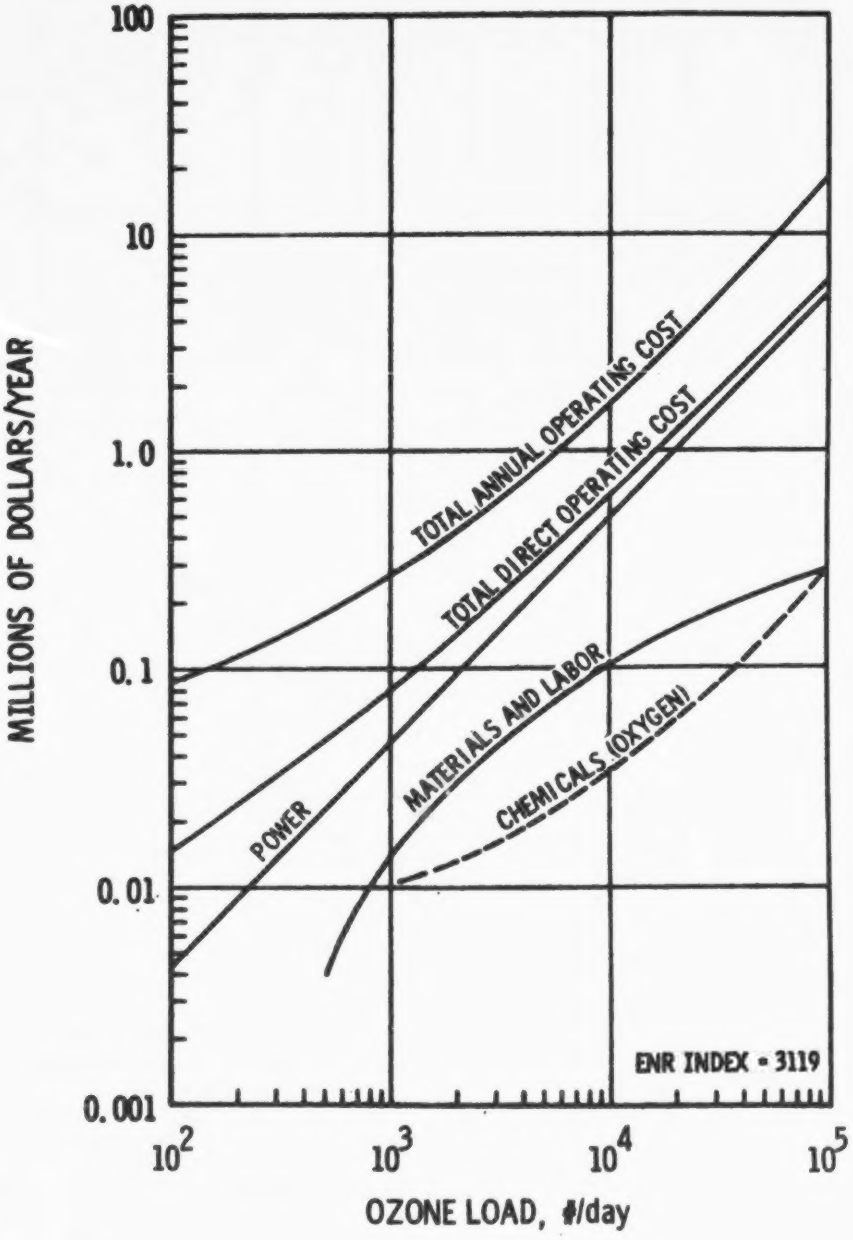
Purchased equipment and installation cost for estimation of the total capital investment includes deaerators, process pump, injector, mixers, reactors, holding tanks, oxygen compressors, dryers, ozone generators, sumps and draws, ozone decomposer, and ozone dosage of 55 mg/L derived from an oxygen feed.

IV.5.14.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.


10,000
1,000
100
10
1.0
0.1
MILLIONS OF DOLLARS
TOTAL CAPITAL INVESTMENT
PURCHASED AND INSTALLED EQUIPMENT
ENR INDEX = 3119
10^2
10^3
10^4
10^5
OZONE LOAD, #/day

ANNUAL OPERATING COST



IV.5.15 CHEMICAL REDUCTION [1]

IV.5.15.1 Description

Chemical reduction is used to reduce metals to less toxic oxidation states. Reduction-oxidation or "redox" reactions are those in which the oxidation state of at least one reactant is raised while that of another is lowered. In the reaction

$$2H_2CrO_4 + 3SO_2 + 3H_2O \rightarrow Cr_2(SO_4)_3 + 5H_2O \qquad (1)$$

the oxidation state of Cr changes from 6^+ to 3^+ (Cr is reduced); the oxidation state of S increases from 2^+ to 3^+ (S is oxidized).. This change of oxidation state implies that an electron was transferred from S to Cr(VI). The decrease in the positive valence (or increase in the negative valence) with reduction takes place simultaneously with oxidation in chemically equivalent ratios. Reduction is used to treat wastes in such a way that the reducing agent lowers the oxidation state of a substance in order to reduce its toxicity, reduce its solubility, or transform it into a form that can be more easily handled.

The base metals are good reducing agents, as evidenced by the use of iron, aluminum, zinc, and sodium compounds for reduction treatments. In addition, sulfur compounds also appear among the more common reducing agents.

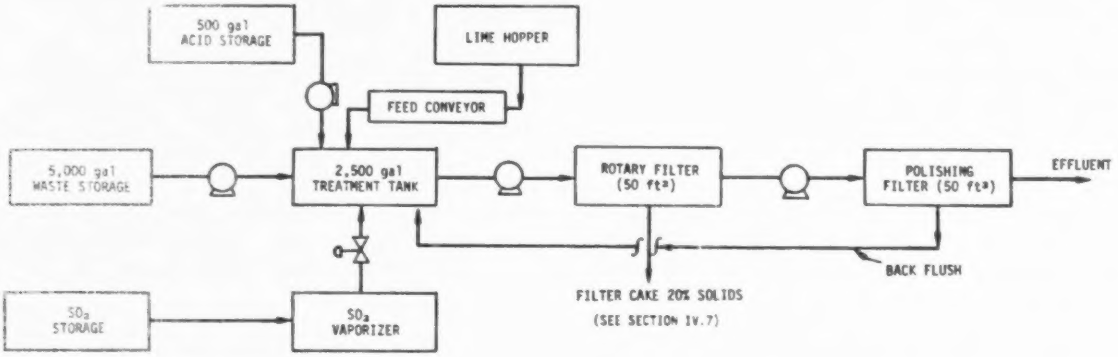
Liquids are the primary waste form treatable by chemical reduction. The most powerful reductants are relatively nonselective; therefore, any easily reducible material in the waste stream will be treated. For example, in reducing heavy metals to remove them from a waste oil, quantities of esters large enough to cause odor problems may also be formed by the reduction.

Gases such as chlorine dioxide and chlorine have been treated by reducing solutions for the small-scale disposal of gas in laboratories. For reduction of fluorine, instead of a solution, a scrubber filled with solid bicarbonate, soda lime or granulated carbon is recommended. Reduction has limited application to slurries, tars, and sludges, because of the difficulties of achieving intimate contact between the reducing agent and the hazardous constituent; consequently, the reduction process would be very inefficient.

In general, hazardous materials occurring as powders or other solids usually have to be solubilized prior to chemical reduction. The first step of the chemical reduction process is usually the adjustment of the pH of the solution to be treated. With sulfur dioxide treatment of chromium (VI), for instance, the reaction requires a pH in the range of 2 to 3. The pH adjustment is done with the appropriate acid (e.g., sulfuric). This is followed by addition of the reducing agent. Mixing is provided to improve contact between the reducing agent and the waste. The agent can be in the form of a gas (sulfur dioxide) or solution (sodium borohydride) or perhaps finely divided powder if there is adequate mixing. Reaction times vary for different wastes, reducing agents, temperatures, pH, and concentration. For commercial-scale operations for treating chromium wastes, reaction times are in the order of minutes. Additional time is usually allowed to ensure complete mixing and reduction. Once reacted, the reduced solution is generally subjected to some form of treatment to settle or precipitate the reduced material. A treatment for the removal of what remains of the reducing agent may be included. This can be unused reducing agent or the reducing agent in its oxidized state. Unused alkali metal hydrides are decomposed by the addition of a small quantity of acid. The pH of the reaction medium is typically increased so that the reduced material will precipitate out of solution. Filters or clarifiers are often used to improve separation.

While some stream components may be added or removed, the output stream from a chemical reduction treatment is not very different from the input stream. Reducing agents, such as sodium borohyride and zinc, introduce to the reaction mixture ions that are not easily separable from the product streams. The effluent solution is typically acidic and must be neutralized prior to discharge with materials such as hydrated lime, caustic soda, or soda ash.

Equipment normally associated with chemical reduction processes includes storage and contact vessels, metering equipment, agitators, and instrumentation for determination of pH and degree of completion of the reduction reaction. A flow diagram of a chemical reduction system is shown below.


500 gal ACID STORAGE
LIME HOPPER
FEED CONVEYOR
5,000 gal WASTE STORAGE
2,500 gal TREATMENT TANK
ROTARY FILTER (50 ft²)
POLISHING FILTER (50 ft²)
EFFLUENT
BACK FLUSH
FILTER CAKE 20% SOLIDS
(SEE SECTION IV.7)
SO₂ STORAGE
SO₂ VAPORIZER

TREATMENT: BATCH

WASTE: CONCENTRATED CHROME WASTE
100,000 ppm CrO_3; 85% as Cr^{+3}
IN 20%H_2SO_4
WASTE PROCESSING CAPACITY: 2,000 gal/shift
OPERATING PERIOD: 240 days/yr
8 hours/day

RAW MATERIALS:

240 lb/day SO_2
2,065 lb/day lime

IV.5.15.2 Typical Design Criteria

The design parameters for a chemical reduction treatment process are dependent on a variety of factors, such as the flow rate, and the utilized reductant. For instance, with sulfur dioxide treatment of chromium (Vl) the reaction requires a pH of 2 to 3. A typical chromium (Vl) treatment process using sulfur dioxide needs 240 lb/day of sulfur dioxide and 2,065 lb/day of lime based on a waste processing capacity of 2,000 gal per shift. In general, contact times range from 1 to 5 minutes.

IV.5.15.3 Costs

The cost of treatment at individual plant locations and on individual wastes varies greatly. Capital costs depend upon such factors as the type, volume, and composition of the waste; degree of treatment required; treatment process selected; availability of required services; and the specific material to be recovered (i.e., metals, chemicals, or water). For the purpose of cost estimating, data are generated for the chemical reduction of chromium waste from a plating operation using sulfur dioxide treatment. A flow diagram of this system and its operating characteristics was presented in Section IV.5.15.1.

Cost item	Annual quantity	Cost per unit quantity	Annual cost, $	
Direct operating cost				
Labor				
Operating	2,500 man-hr	$16/hr	40,000	
Maintenance			4,700	
				44,700
Chemicals				
Sulfur dioxide	29 tons	$158/ton	4,600	
Lime	250 tons	$33.5/ton	8,400	
				13,000
Power	35,000 kWh	$0.035/kWh		1,200
Materials				4,900
Total				63,800
Total indirect operating cost				84,500
Total annual operating cost				193,000

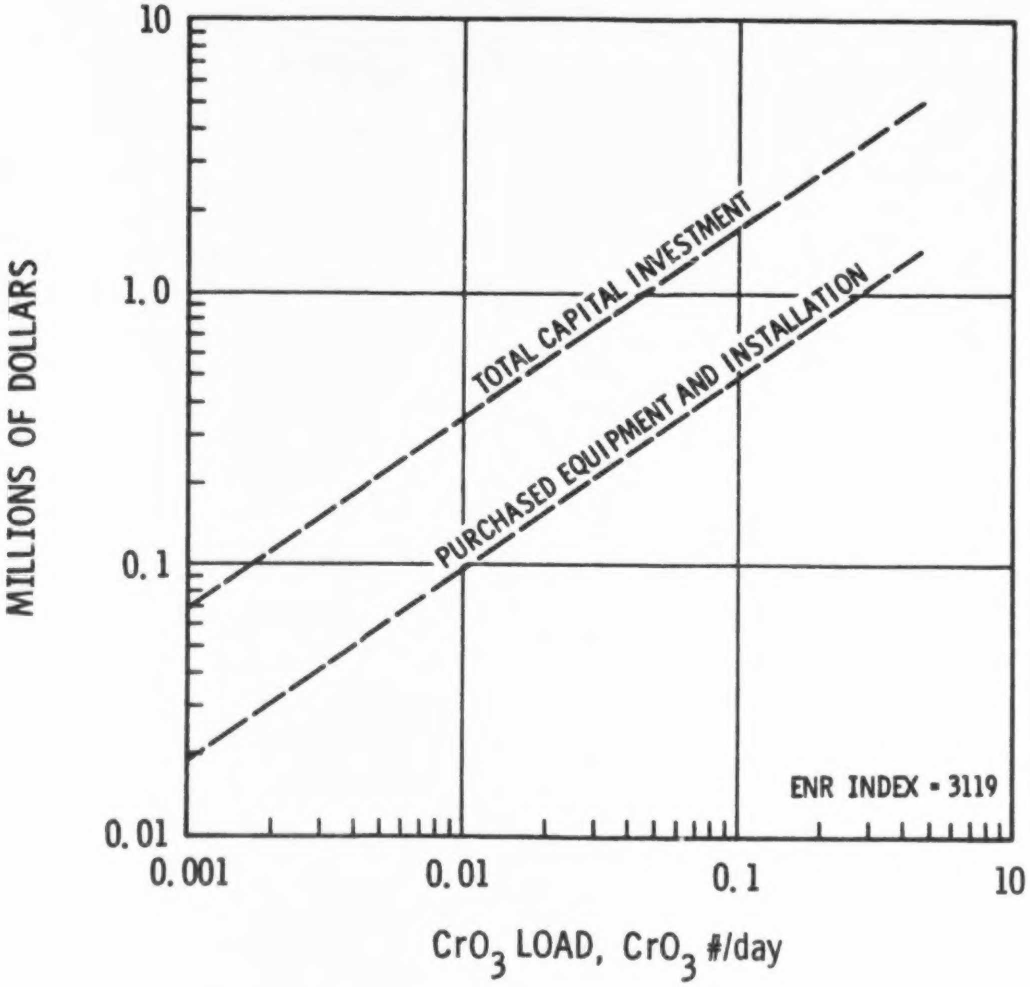
Cost curves were developed for the total capital investment and annual operating cost using the cost data shown on the previous page and the following exponential scaling factors, which were used to determine costs at varying flowrates:

Cost item	Exponential factor
Total capital investment	0.7
Labor	0.3
Power and chemicals	1.0

IV.5.15.4 References

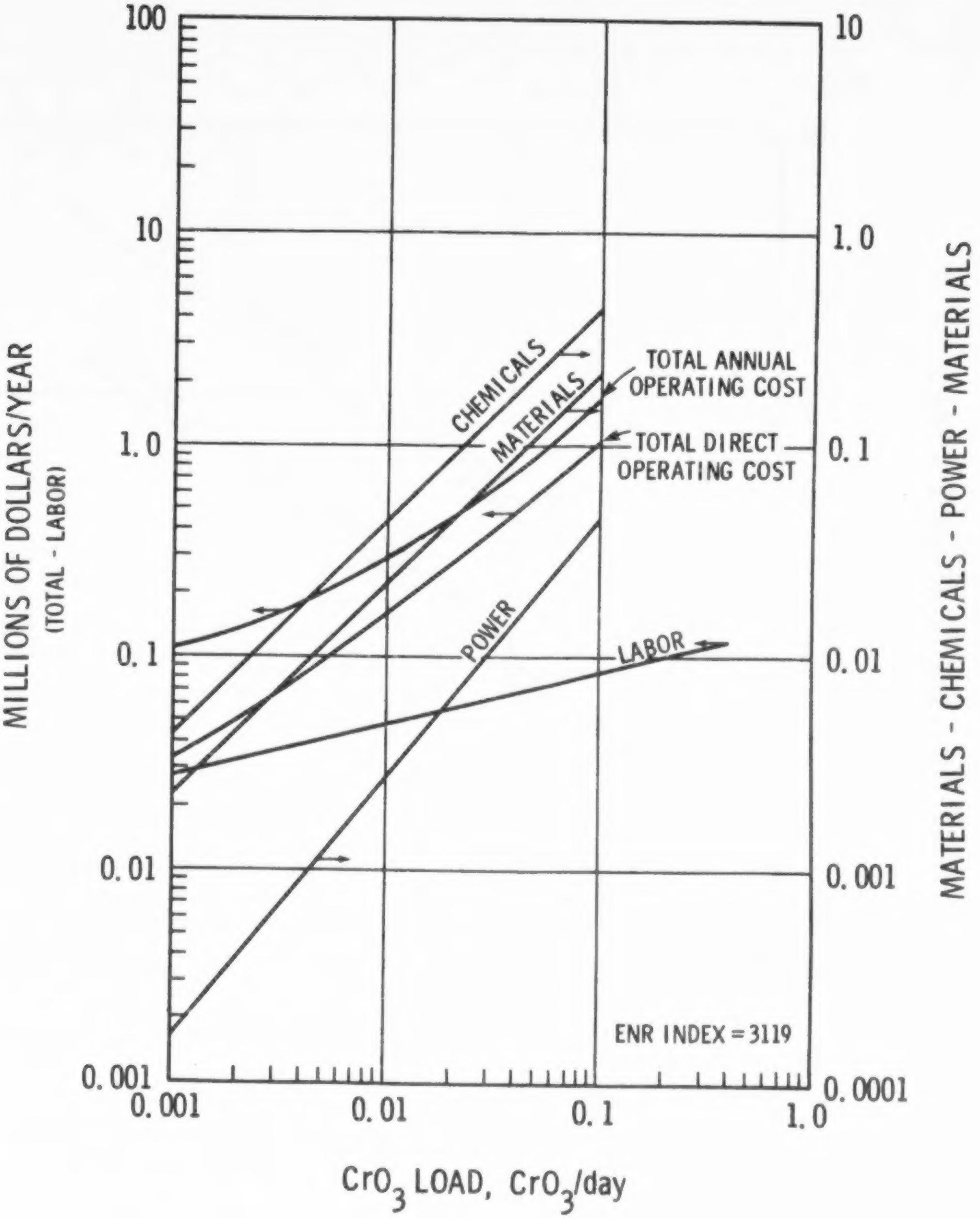
1. Physical, Chemical, and Biological Treatment Techniques for Industrial Wastes, PB 275 287, U.S. Environmental Protection Agency, Washington, D.C., November 1976. pp. 38-1 to 38-13.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



IV.6 SLUDGE TREATMENT

IV.6.1 GRAVITY THICKENING [1]

IV.6.1.1 Description

Thickening of sludge consists of the removal of supernatant, thereby reducing the volume of sludge that requires disposal or further treatment. Gravity thickening takes advantage of the difference in specific gravity between the solids and water.

A gravity thickener normally consists of two truss-type steel scraper arms mounted on a hollow pipe shaft keyed to a motorized hoist mechanism. A truss-type bridge is fastened to the tank walls or to steel or concrete columns. The bridge spans the tank and supports the entire mechanism. The thickener resembles a conventional circular clarifier with the exception of having a greater bottom slope.

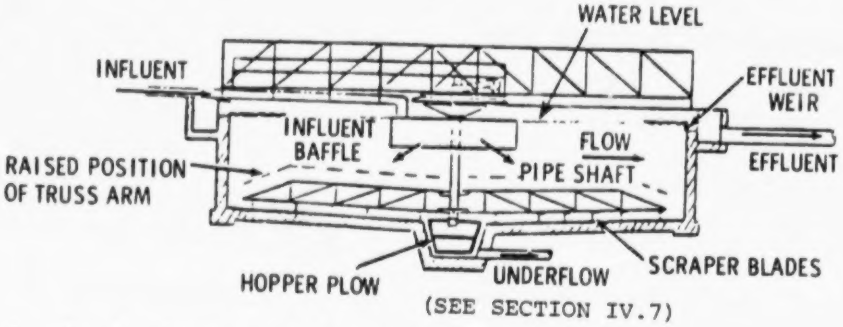
Sludge enters at the middle of the thickener, and the solids settle into a sludge blanket at the bottom. The concentrated sludge is very gently agitated by the moving rake, which dislodges gas bubbles and prevents bridging of the sludge solids. It also keeps the sludge moving toward the center well from which it is removed. Supernatant liquor passes over an effluent weir around the circumference of the thickener. In the operation of gravity thickeners, it is desirable to keep a sufficiently high flow of fresh liquid entering the concentrator to prevent the development of septic conditions and resulting odors.

Gravity thickening is characterized by zone settling. The four basic settling zones in a thickener are:

- The clarification zone at the top containing the relatively clear supernatant
- The hindered settling zone where the suspension moves downward at a constant rate and a layer of settled solids begins building from the bottom of the zone.
- The transition zone characterized by a decreasing solids settling rate
- The compression zone where consolidation of sludge results solely from liquid being forced upward around the solids

Gravity thickener tanks can be square or round; the round variety is much more prevalent. Tanks can be manufactured of concrete or steel. Chemicals can be added to aid in the sludge dewatering.

Equipment normally associated with gravity thickening systems includes sedimentation units and chemical feed equipment. A flow diagram of a gravity thickener is shown below.


WATER LEVEL
INFLUENT
EFFLUENT WEIR
INFLUENT BAFFLE
FLOW
RAISED POSITION OF TRUSS ARM
PIPE SHAFT
EFFLUENT
HOPPER PLOW
UNDERFLOW
(SEE SECTION IV.7)
SCRAPER BLADES

IV.6.1.2 Typical Design Criteria

Detention times of one to three days are generally used. Sludge blankets at least three feet deep are common. Side water depths are at least ten feet in general practice. The following tabulation represents a performance summary with no chemical conditioning.

Type of sludge	Solids surface loading, lb/d/ft²	Thickened sludge solids concentration, %
Primary	20 - 30	8 - 10
Waste activated	5 - 6	2.5 - 3
Trickling filter	8 - 10	7 - 9
Limed tertiary	60	12 - 15
Primary and activated	6 - 10	4 - 7
Primary and trickling filter	10 - 12	7 - 9
Limed primary	20 - 25	7 - 12

IV.6.1.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes the thickener and all related mechanical equipment. Pumps are not included. Total annual operating costs do not include polymer addition. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Thickening of secondary sludge	820 lb/Mgal
Loading	6 $\text{lb/ft}^2\text{/d}$

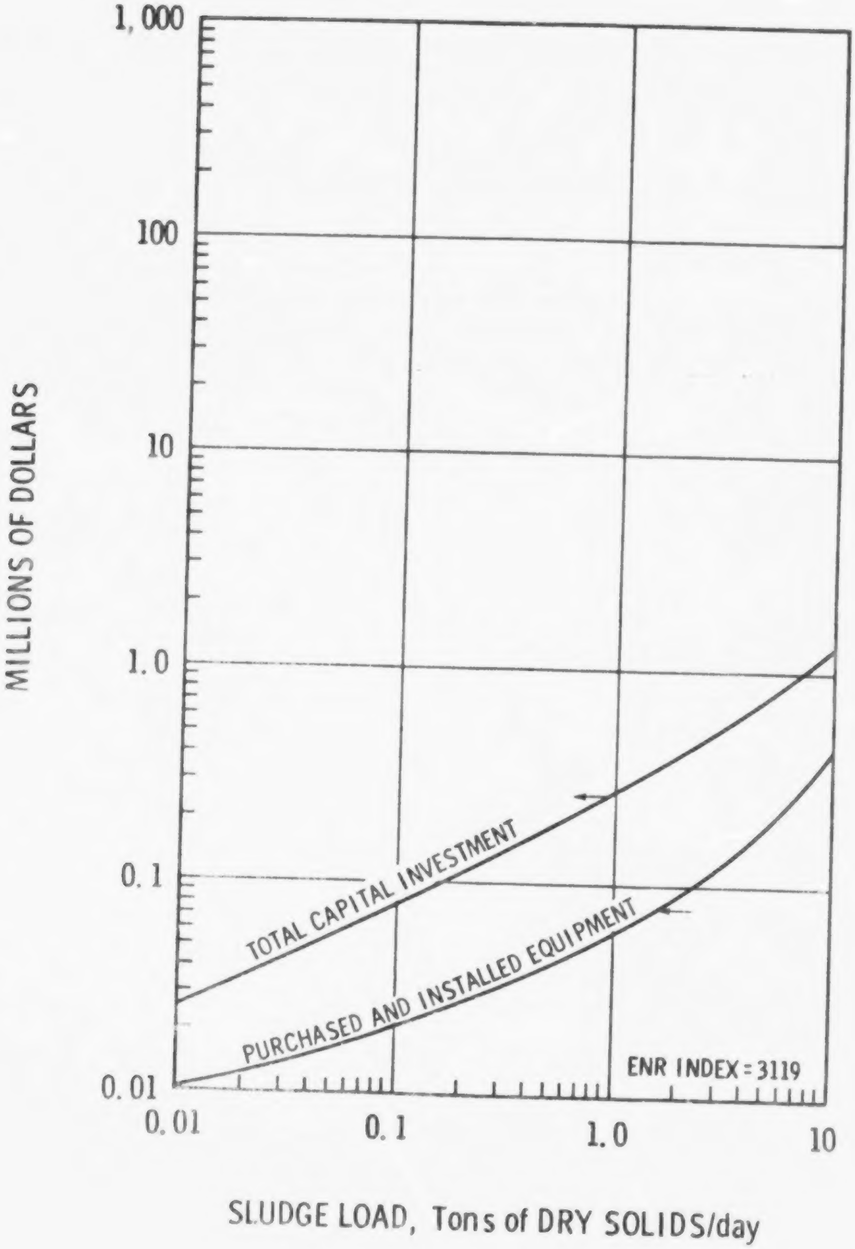
To adjust costs for alternative sludge quantities, concentrations, and thickening properties, calculate the effective sludge load, L_e, using the following equation:

$$L_e = L_{Design} \times \frac{6\ \text{lb/ft}^2\text{/d}}{\text{new design mass loading}} \times \frac{\text{new design sludge mass}}{820\ \text{lb/Mgal}}$$

IV.6.1.4 References

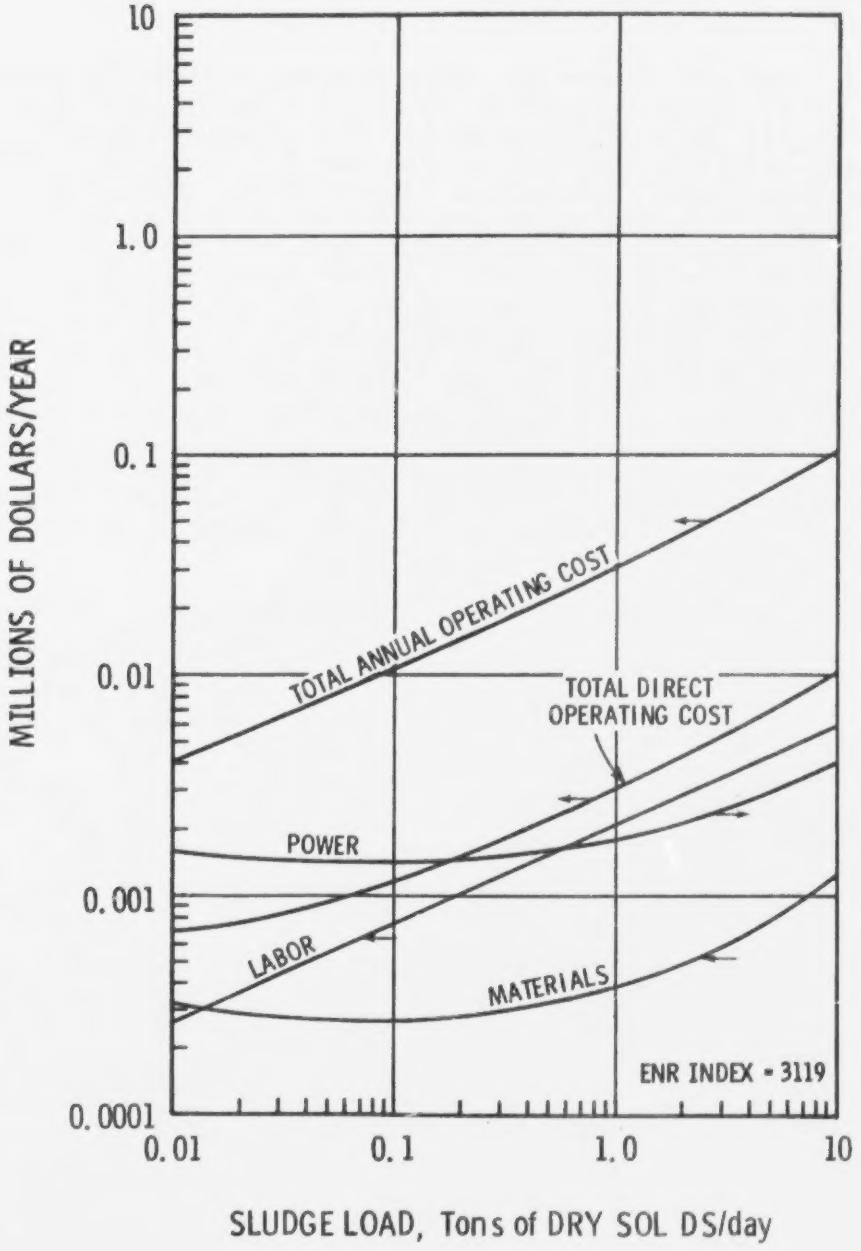
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



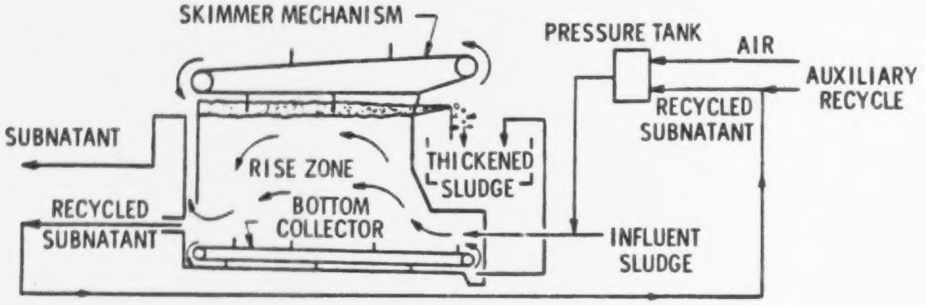
IV.6.2 FLOTATION THICKENING [1]

IV.6.2.1 Description

Flotation (dissolved air flotation) thickening utilizes air to float sludge to the surface of the thickener, thereby reducing the water content and volume of the sludge.

In a dissolved air flotation (DAF) system, a recycled subnatant flow is pressurized from 30 to 70 lb/in.^{2} (gage) and then saturated with air in a pressure tank. The pressurized effluent is then mixed with the influent sludge and subsequently released into the flotation tank. The excess dissolved air then separates from solution, which is now under atmospheric pressure, and the minute (average diameter 80 μm) rising gas bubbles attach themselves to particles that form the floating sludge blanket. The thickened blanket is skimmed off and pumped to the downstream sludge handling facilities while the subnatant is returned to the plant. Polyelectrolytes are frequently used as flotation aids to enhance performance and create a thicker sludge blanket. A description of the DAF process in general is presented in Section IV.3.4.

Equipment normally associated with a flotation thickening system includes dissolved air flotation units and air compressors. A flow diagram for a flotation thickening system is shown below.


SKIMMER MECHANISM
PRESSURE TANK
AIR
AUXILIARY RECYCLE
RECYCLED SUBNATANT
SUBNATANT
RISE ZONE
THICKENED SLUDGE
RECYCLED SUBNATANT
BOTTOM COLLECTOR
INFLUENT SLUDGE

IV.6.2.2 Typical Design Criteria

Typical design criteria for flotation thickening are shown on the next page.

Design criteria	Value
Pressure	30 - 70 lb/in.^{2} gage
Effluent recycle ratio	30 - 150% of influent flow
Air-to-solids ratio	0.005 -0.06 lb air/lb solids
Solids loading	5 - 55 lb/ft^2/d (depending on sludge type and whether flotation aids are used)
Polyelectrolyte addition (when used)	5 - 10 lb/ton of dry solids
Solids capture	70 - 98$^+$%
Total solids in unthickened sludge	0.3 - 2.0%
Total solids in thickened sludge	3 - 12%
Hydraulic loading	0.4 - 2.0 gpm/ft^2

Conditions for specific sludge types are shown below.

Sludge type	Feed solids concentration, %	Typical loading rate without polymer, lb/ft^2/d	Typical loading rate with polymer, lb/ft^2/d	Float solids concentration, %
Primary + WAS[a]	2.0	20	60	5.5
Primary + (WAS + $FeCl_3$)	1.5	15	45	3.5
(Primary + $FeCl_3$) + WAS	1.8	15	45	4.0
WAS	1.0	10	30	3.0
WAS + $FeCl_3$	1.0	10	30	2.5
Digested primary + WAS	4.0	20	60	10.0
Digested primary + (WAS + $FeCl_3$)	4.0	15	45	8.0
Tertiary, alum	1.0	8	24	2.0

[a] Waste activated sludge.

IV.6.2.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes the flotation chamber (2 hr detention time based on sludge flow), pressure tanks (60 lb/in.^{2} gage), and recycle pumps (100% recycle). The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Thickening of secondary sludge	820 lb/Mgal
Loading rate	2 lb/ft^2/hr
Operating hours	0.1 and 1 Mgal/d = 40 hr/wk 10 Mgal/d = 100 hr/wk; 100 Mgal/d = 168 hr/wk

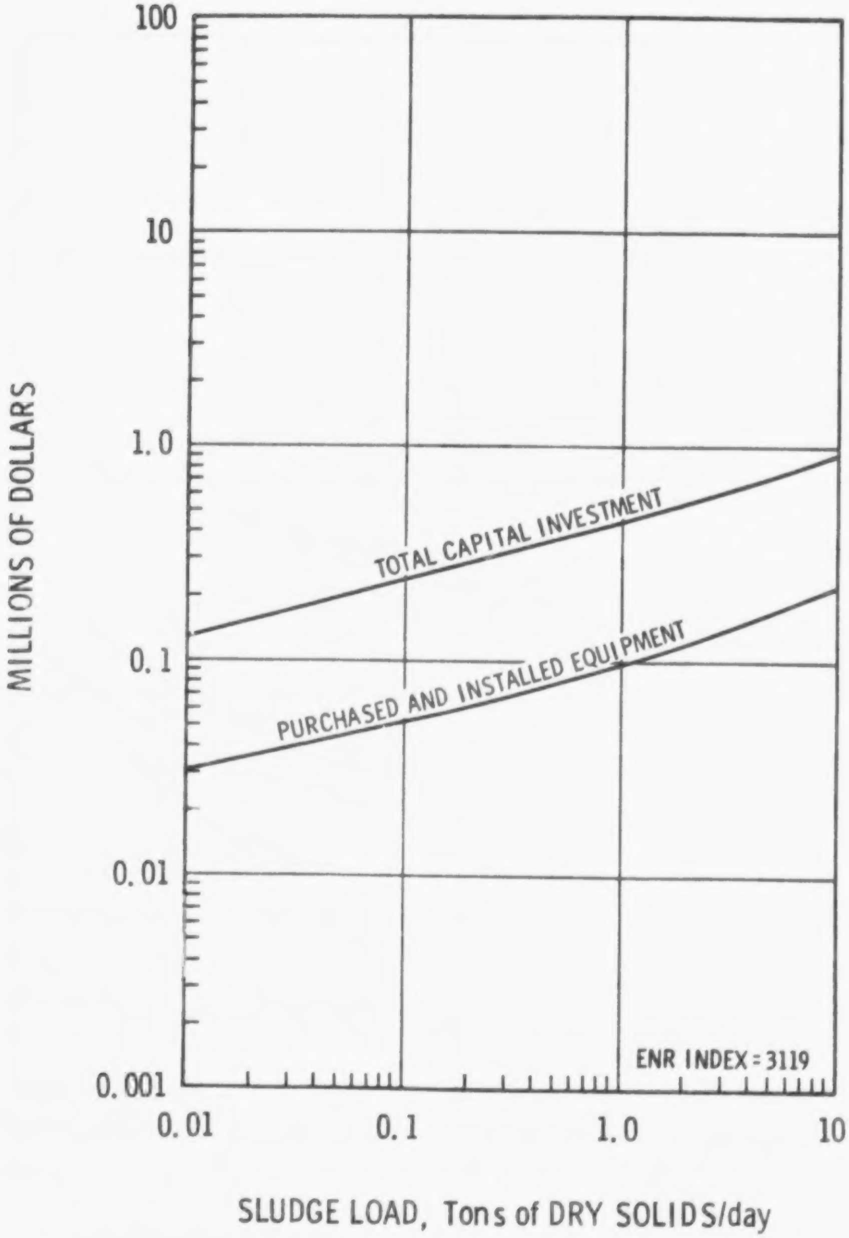
To determine costs at loading rates or sludge quantities other than those specified on the previous page, calculate the effective flow using the following equation:

$$Q_E = Q_{Design} \times \frac{2\ \text{lb/ft}^2\text{/hr}}{\text{new design mass loading rate}} \times \frac{\text{new design sludge mass}}{820\ \text{lb/d/Mgal}}$$

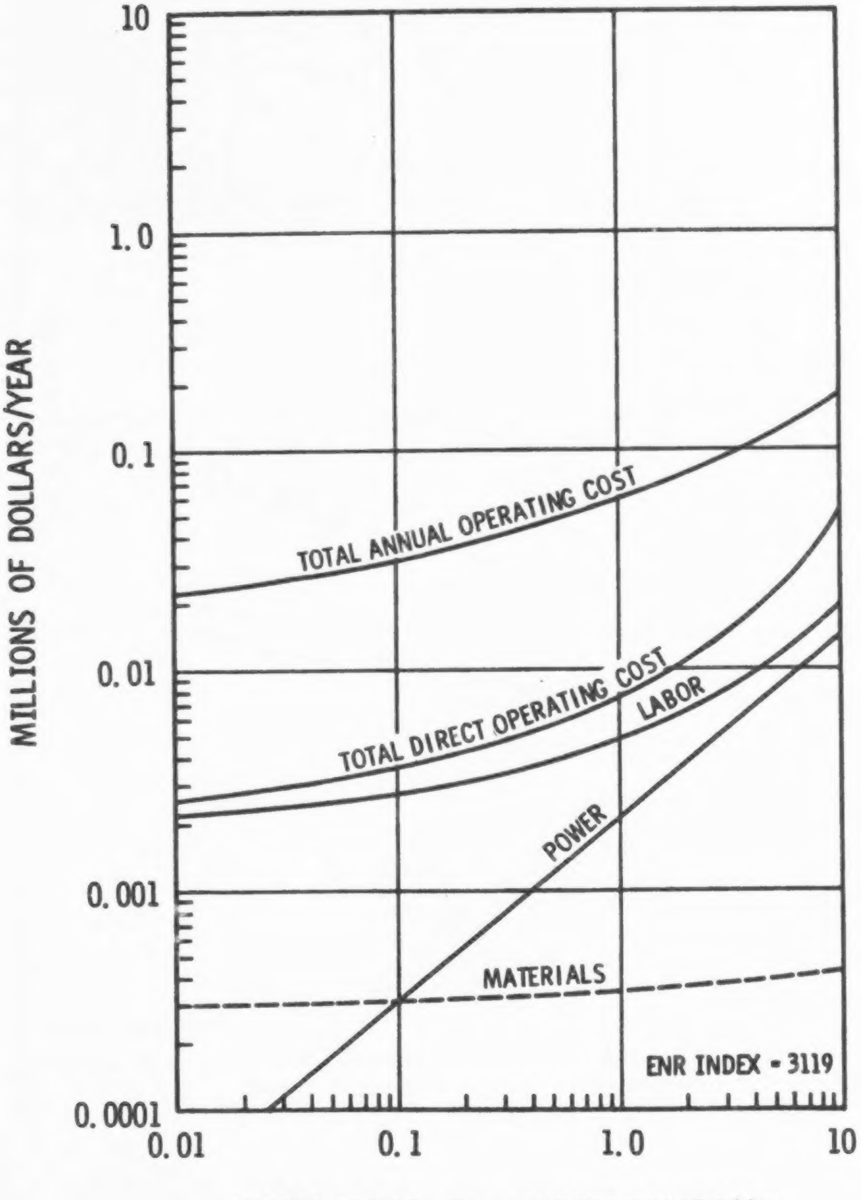
IV.6.2.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



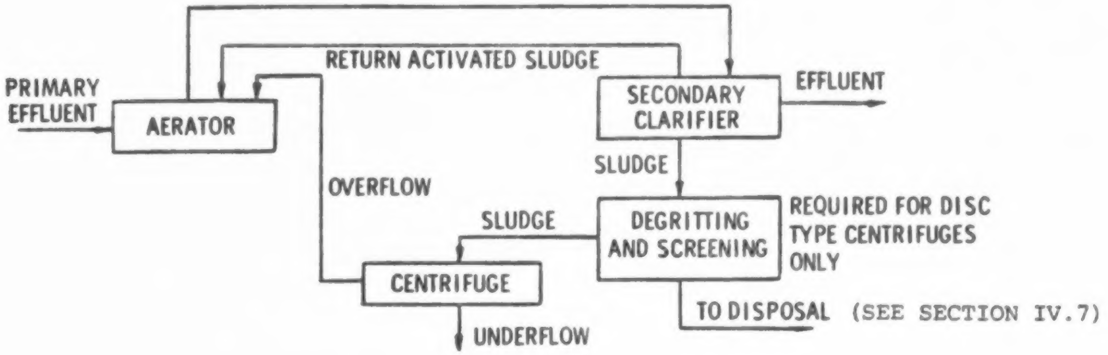
IV.6.3 CENTRIFUGAL THICKENING [1]

IV.6.3.1 Description

Centrifugal thickening is the thickening of sludges using disc, basket, or solid-bowl centrifuges.

Centrifuges may be used to thicken sludges by the use of centrifugal force to increase the sedimentation rate of sludge solids. The three most common types of units are the continuous solid-bowl type, the disc type, and the basket type. Refer to Section IV.6.12 for unit descriptions.

Equipment normally associated with a centrifugal thickening systems includes the centrifuge, sludge feed pump, solids conveyor, and concentrate pumps. A flow diagram of a centrifugal thickening system is shown below.


RETURN ACTIVATED SLUDGE
PRIMARY EFFLUENT
AERATOR
SECONDARY CLARIFIER
EFFLUENT
SLUDGE
OVERFLOW
SLUDGE
DEGRITTING AND SCREENING
REQUIRED FOR DISC TYPE CENTRIFUGES ONLY
CENTRIFUGE
TO DISPOSAL (SEE SECTION IV.7)
UNDERFLOW

IV.6.3.2 Typical Design Criteria

Installation of centrifugal thickening systems is site-specific and dependent upon a manufacturer's product line. Maximum capacities of about 100 tons/hr of dry solids are available in solid-bowl units with diameters up to 54 in. and power requirements up to 175 hp. Disc units are also available with capacities up to 400 gpm of concentrate.

IV.6.3.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes the centrifuge, drive motor, screens, strainer, pumps, and degritting system for a disc centrifuge. For a solid-bowl centrifuge unit, these costs include the centrifuge and drive motor only. All units are designed to handle 820 lb/Mgal of dry solids at 0.8% concentration. Polymer costs are not included. Labor costs were assumed at $15/man-hour.

The total capital investment and total annual direct operating costs for disc and solid-bowl centrifuges are shown below based on an ENR Index of 3119. The following cost estimation was based on a disc and solid-bowl type with a flow rate of 300-500 gpm.

Centrifuge type	Influent sludge flow, gpm	Total capital investment, $	Annual operating and maintenance cost, $			Total annual direct operating cost, $
			Power	Material	Labor	
Disc	100 - 150	457,000	11,000	2,300	13,000	26,300
	150 - 300	687,000	22,000	3,500	13,000	38,500
	300 - 550	914,000	40,400	6,900	13,000	60,300
Solid-bowl	50 - 60	413,000	6,000	2,100	9,200	17,300
	up to 150	794,000	14,900	4,000	9,200	28,100
	up to 400	1,484,000	39,700	7,500	9,200	56,400

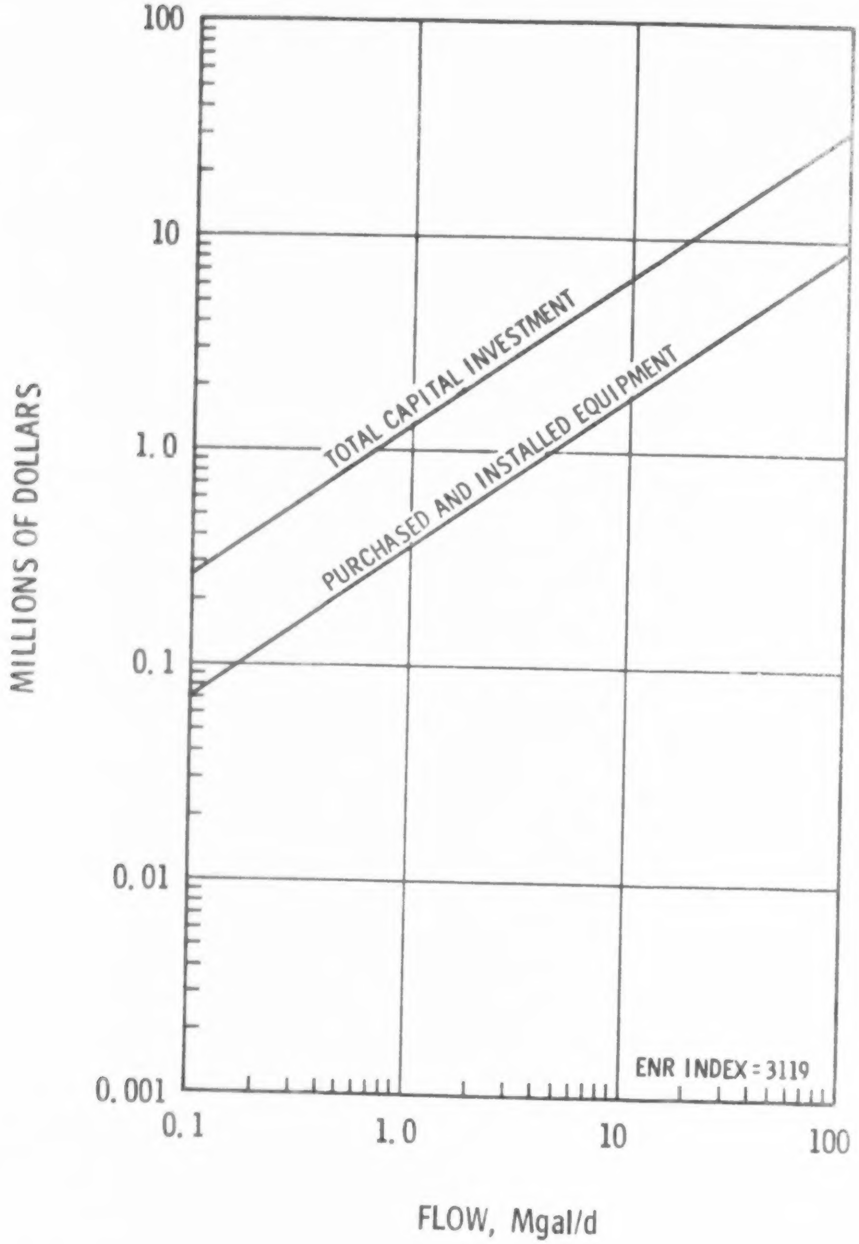
Power cost is based on maximum influent sludge flow return. Only centrifuge drive requirements are considered.

IV.6.3.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT

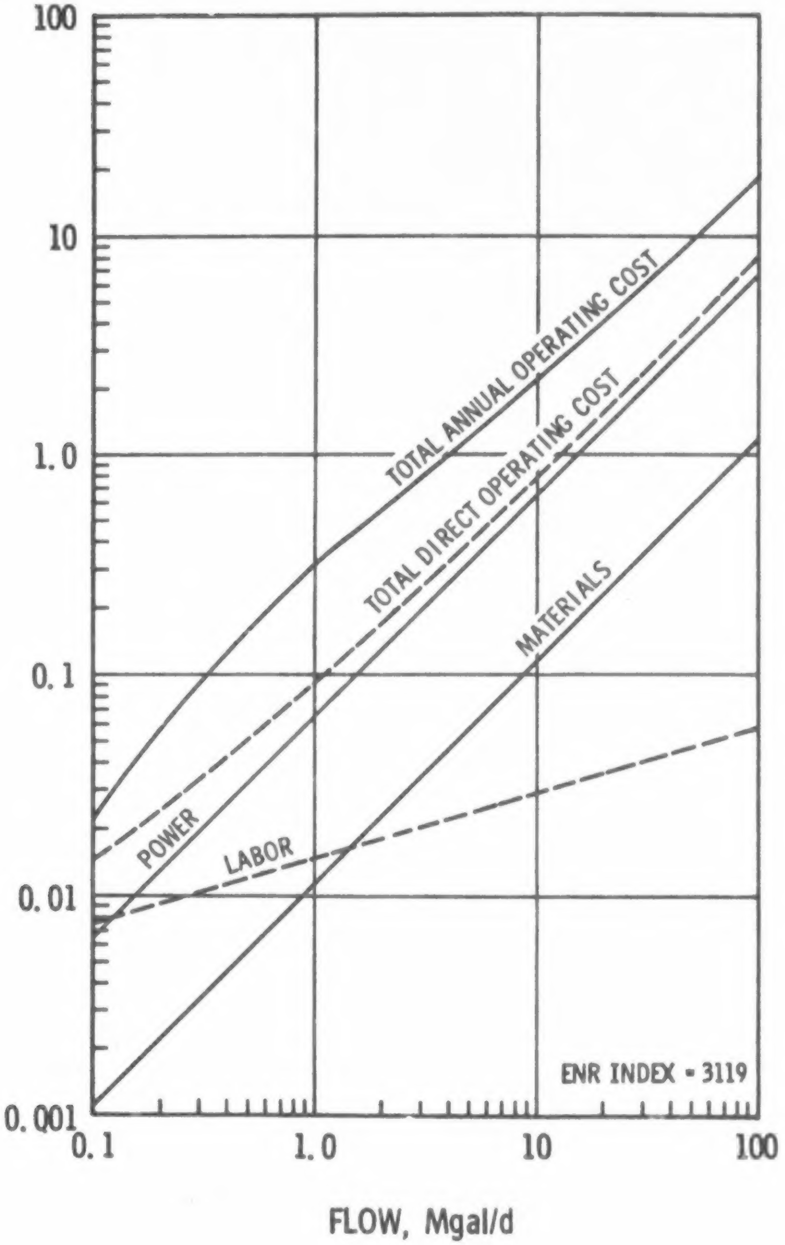
(DISC)



Date: 6/23/80

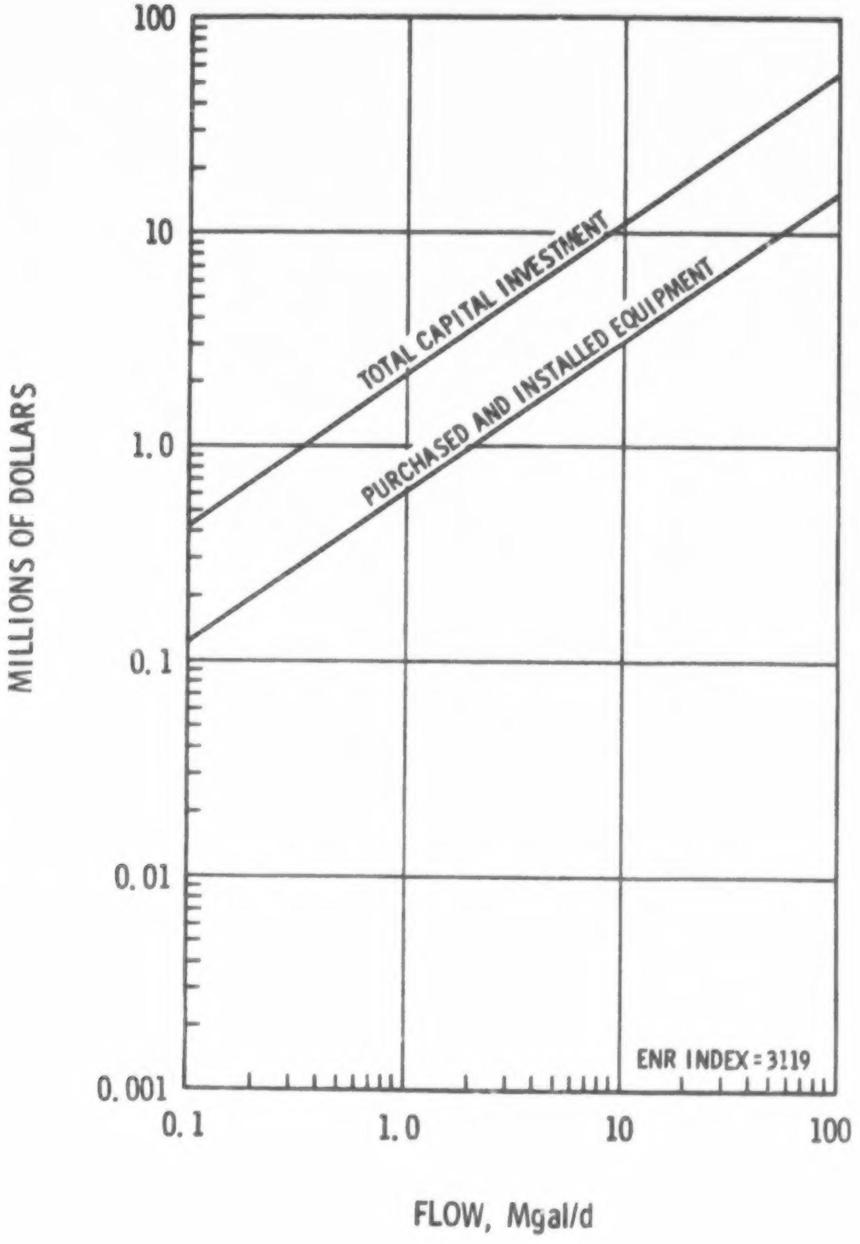
ANNUAL OPERATING COST

(DISC)



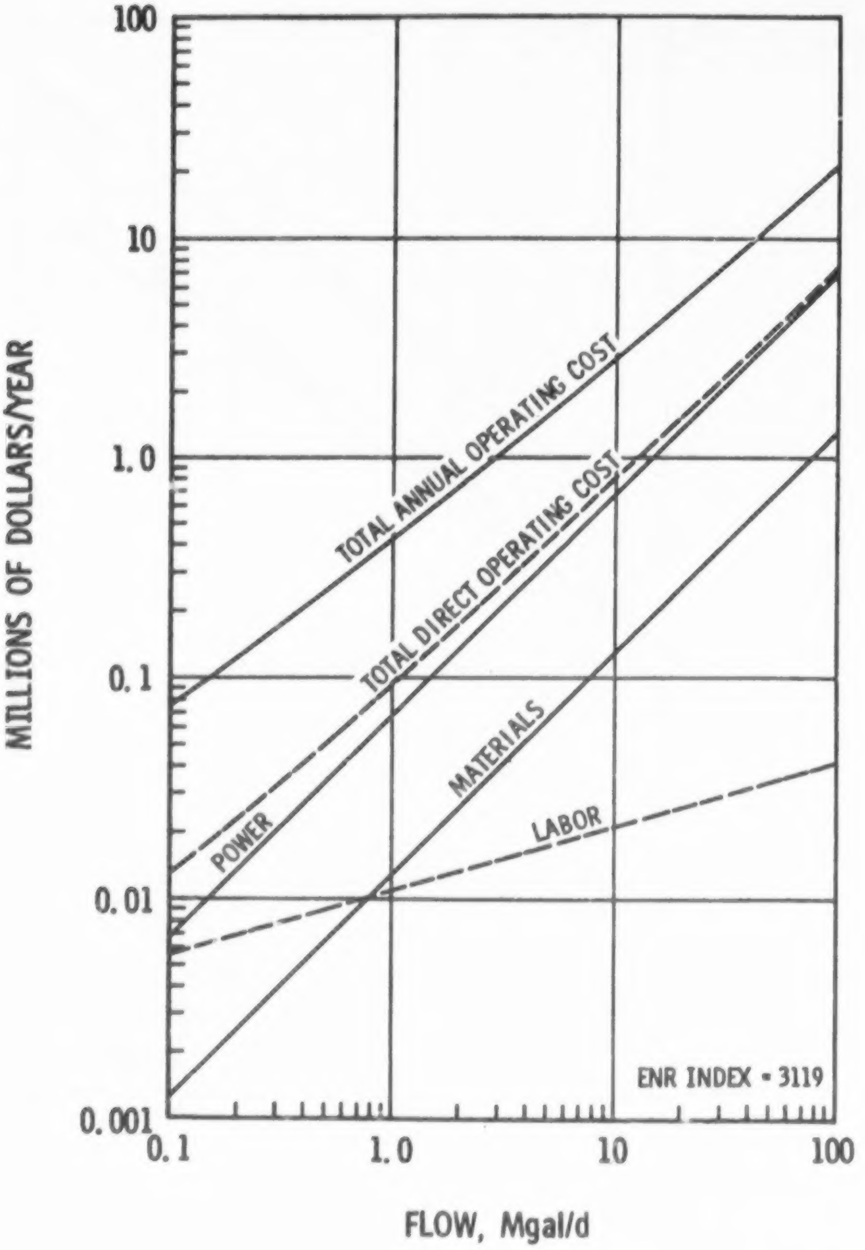
TOTAL CAPITAL INVESTMENT

(SOLID BOWL)



ANNUAL OPERATING COST

(SOLID BOWL)



IV.6.4 AEROBIC DIGESTION [1]

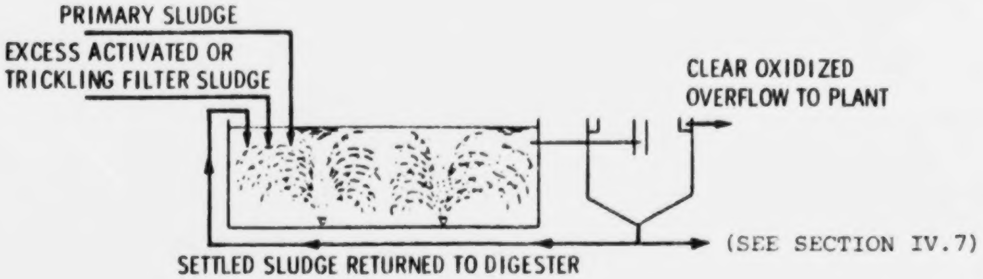
IV.6.4.1 Description

Aerobic digestion is a method of sludge stabilization in an open tank that can be regarded as a modification of the activated sludge process. In the aerobic digestion system, microbiological activity beyond cell synthesis is stimulated by aeration, oxidizing both the biodegradable organic matter and some cellular material into CO_2, H_2O, and NO_3. The oxidation of cellular matter is called endogenous respiration and is normally the predominant reaction occurring in aerobic digestion. Stabilization is not complete until there has been an extended period of primarily endogenous respiration (typically 15 to 20 days).

Major objectives of aerobic digestion include odor reduction, reduction of biodegradable solids, and improved sludge dewaterability. Aerobic bacteria stabilize the sludge more rapidly than anaerobic bacteria, although a less complete breakdown of cells is usually achieved. Oxygen can be supplied by surface aerators or by diffusers. Other equipment may include sludge recirculation pumps and piping, mixers, and scum collection baffles. Aerobic digestors are designed similar to rectangular aeration tanks and use conventional aeration systems, or employ circular tanks and use an eductor tube for deep tank aeration.

Both one- and two-tank systems are used. Small plants often use a one-tank batch system with a complete mix cycle followed by settling and decanting (to help thicken the sludge). Larger plants may consider a separate sedimentation tank to allow continuous flow and facilitate decanting and thickening. Air may be replaced with oxygen.

Typical equipment includes facilities for sludge handling, sludge control, aeration, as well as pumps and mixers. A flow diagram of an aerobic digestion system is shown below.


PRIMARY SLUDGE
EXCESS ACTIVATED OR TRICKLING FILTER SLUDGE
CLEAR OXIDIZED OVERFLOW TO PLANT
(SEE SECTION IV.7)
SETTLED SLUDGE RETURNED TO DIGESTER

IV.6.4.2 Typical Design Criteria

Typical design criteria for an aerobic digestion system are listed below.

Design criteria	Value
Solids retention time for 40% reduction in volatile suspended solids (VSS)	18 - 20 d for mixed sludges from activated sludge or trickling filter plant; 10 - 16 d for waste activated sludge only; 16 - 18 d for activated sludge from plants without primary settling
Volume allowance	3 - 4 ft^3/capita
VSS loading	0.02 - 0.4 $lb/ft^3/d$
Air requirements	20 - 60 ft^3/min/1,000 ft^3
Minimum dissolved oxygen	1 - 2 mg/L
Energy for mechanical mixing	0.75 - 1.25 hp/1,000 ft^3
Oxygen requirements	2 lb/lb of cell tissue destroyed (includes nitrification demand); 1.6 - 1.9 lb/lb of BOD removed in primary sludge

IV.6.4.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes basins (20-day detention time), sludge flow at 5,700 gal/Mgal (1,900 lb/Mgal at 4% solids), and floating mechanical aerators. In addition, power was determined using 1.6 lb O_2/lb VSS destroyed (not including nitrification), and 40% VSS destruction with mechanical aerators at 1.5 lb O_2/hp 1V.

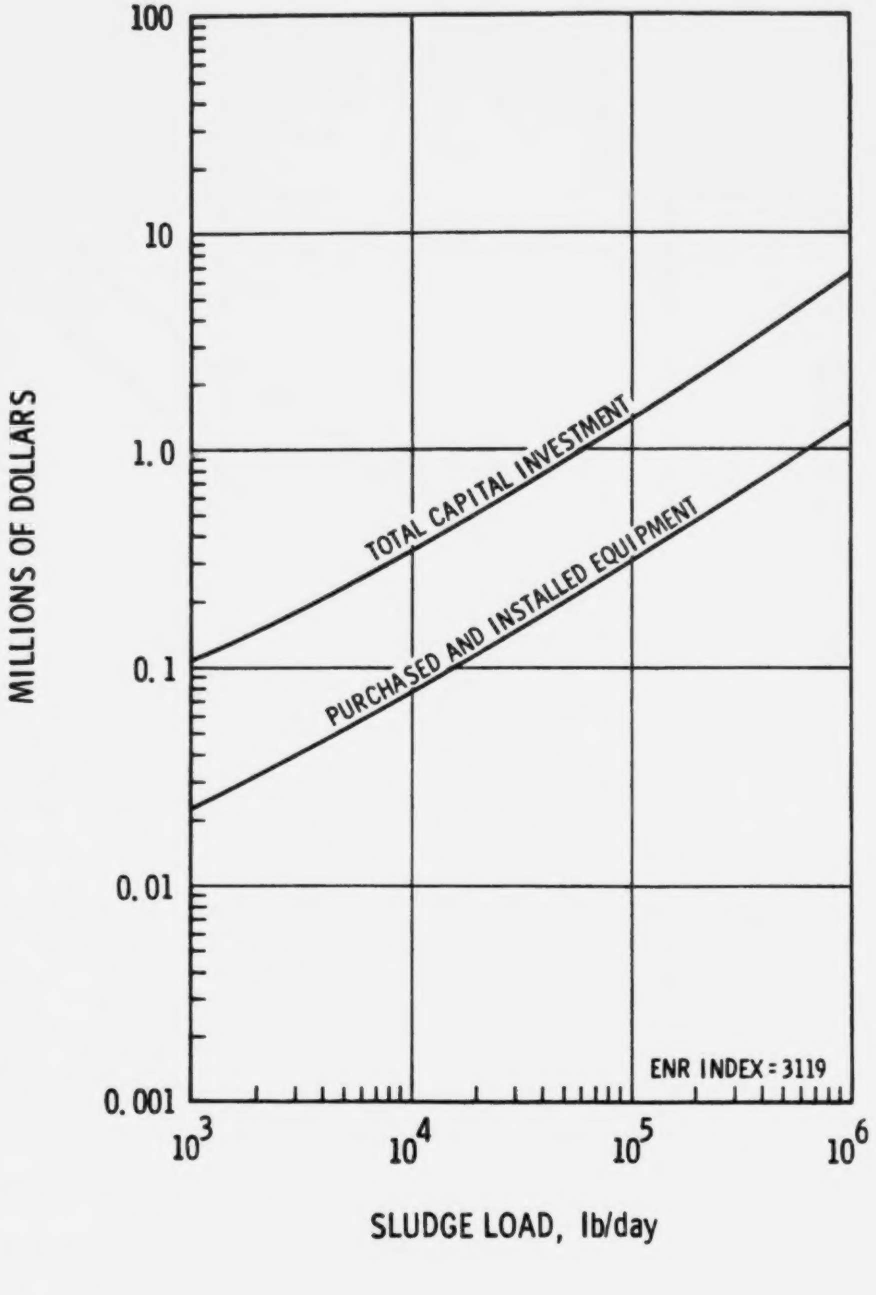
To adjust costs for design factors different from those above, calculate the effective flowrate (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{\text{new design retention time}}{\text{20 days}} \times \frac{\text{new design sludge mass}}{\text{1,900 lb/Mgal}} \times \frac{4\%}{\text{new design sludge concentration}}$$

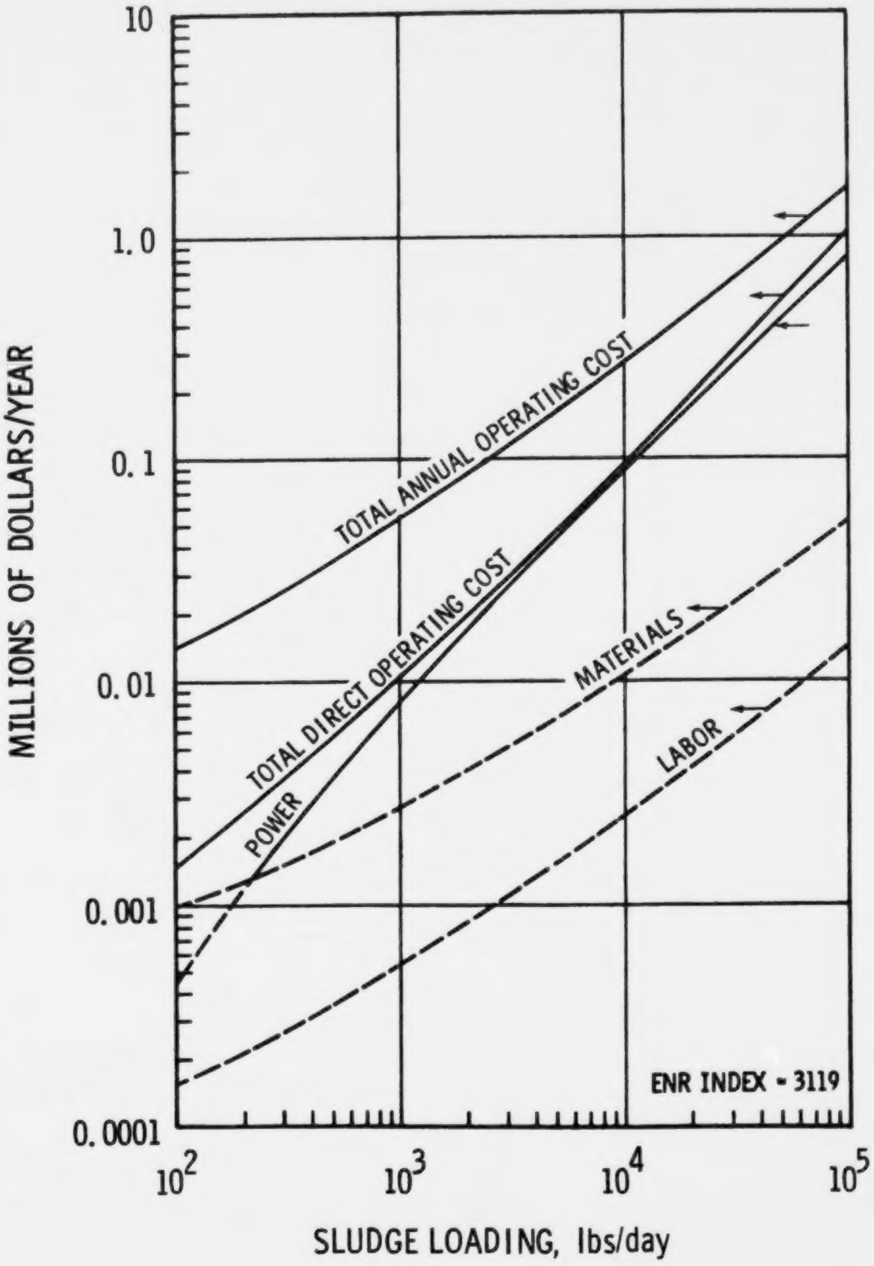
IV.6.4.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.5 ANAEROBIC DIGESTION [1]

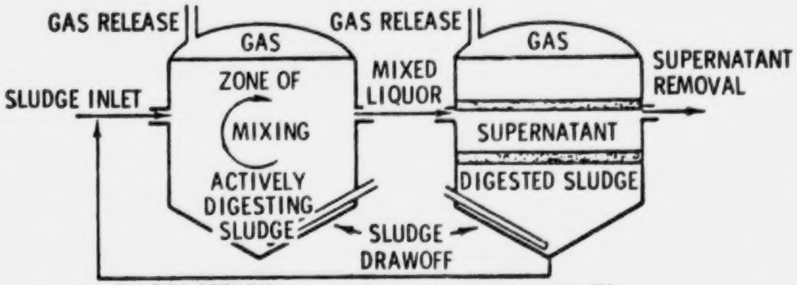
IV.6.5.1 Description

Anaerobic digestion is a process for the breakdown of sludge into methane, carbon dioxide, unusable intermediate organics, and a relatively small amount of cellular protoplasm.

The anaerobic digestion system consists of two vessels that are used for sludge stabilization. The first tank, used for digestion, is equipped with one or more of the following: heater, sludge recirculation pumps, methane gas recirculation, mixers, and scum breaking mechanisms. The second tank is used to store and concentrate the digested sludge and to form a supernatant.

The anaerobic digestion process consists of two distinct simultaneous stages of conversion of organic material by acid-forming bacteria and gasification of the organic acids by methane-forming bacteria. The methane-producing bacteria are very sensitive to conditions of their environment and require careful control of temperature, pH, excess concentrations of soluble salts, metal cations, oxidizing compounds, and volatile acids. They also show an extreme substrate specificity. The digester requires periodic cleanout (from 1 to 2 years) due to buildup of sand and gravel on the digester bottom.

Typical equipment includes sludge handling and control units, pumps, heating equipment, digestion tank equipment, and gas holders. A flow diagram of the two-stage anaerobic digestion system is shown below.


GAS RELEASE
GAS
GAS RELEASE
GAS
SUPERNATANT REMOVAL
ZONE OF
MIXED LIQUOR
SLUDGE INLET
MIXING
SUPERNATANT
ACTIVELY DIGESTING SLUDGE
DIGESTED SLUDGE
SLUDGE DRAWOFF
SLUDGE RETURN

(SEE SECTION IV.7)

IV.6.5.2 Typical Design Criteria

Typical design criteria for anaerobic digestion are shown on the next page.

Design criteria	Value
Solids retention time (SRT)	Depends on temperature; SRT in mesophitic range include 55 d @ 50°F; 40 d @ 67°F; 30 d @ 75°F; 25 d @ 85°F; and 20 d @ 95°F
Tank size	20 - 115 ft diameter; 25 - 45 ft depth; bottom vertical slope = 1 vertical/4 horizontal
Solids loading	0.04 - 0.40 lb VSS/ft³/d
pH	6.7 - 7.6

IV.6.5.3 Costs

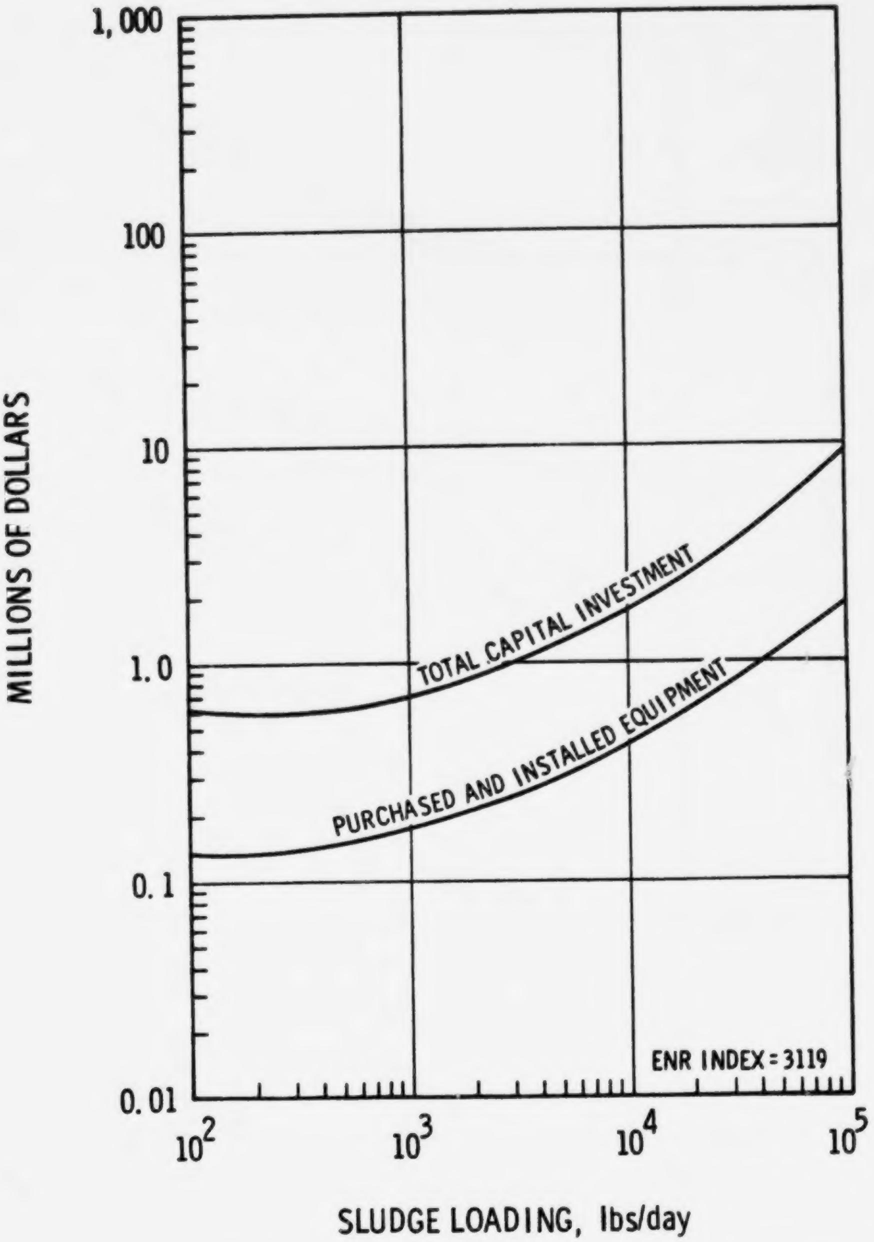
Purchased equipment and installation cost for estimation of the total capital investment includes the digester, heat exchanger, gas collection equipment, and control building. The following operating characteristics were assumed for cost estimation.

Operating characteristics	Assumed value
Service life	50 yr
Feed to digester (primary and secondary are combined)	1,900 lb/Mgal at 4% solids (75% volatile)
Effluent from digesters	900 lb/Mgal at 2.5% solids
Loading rate	0.16 lb/ft³/d
Operating temperature	85 - 110°F
Gas utilization	Digester gas utilized for heating; excess not utilized

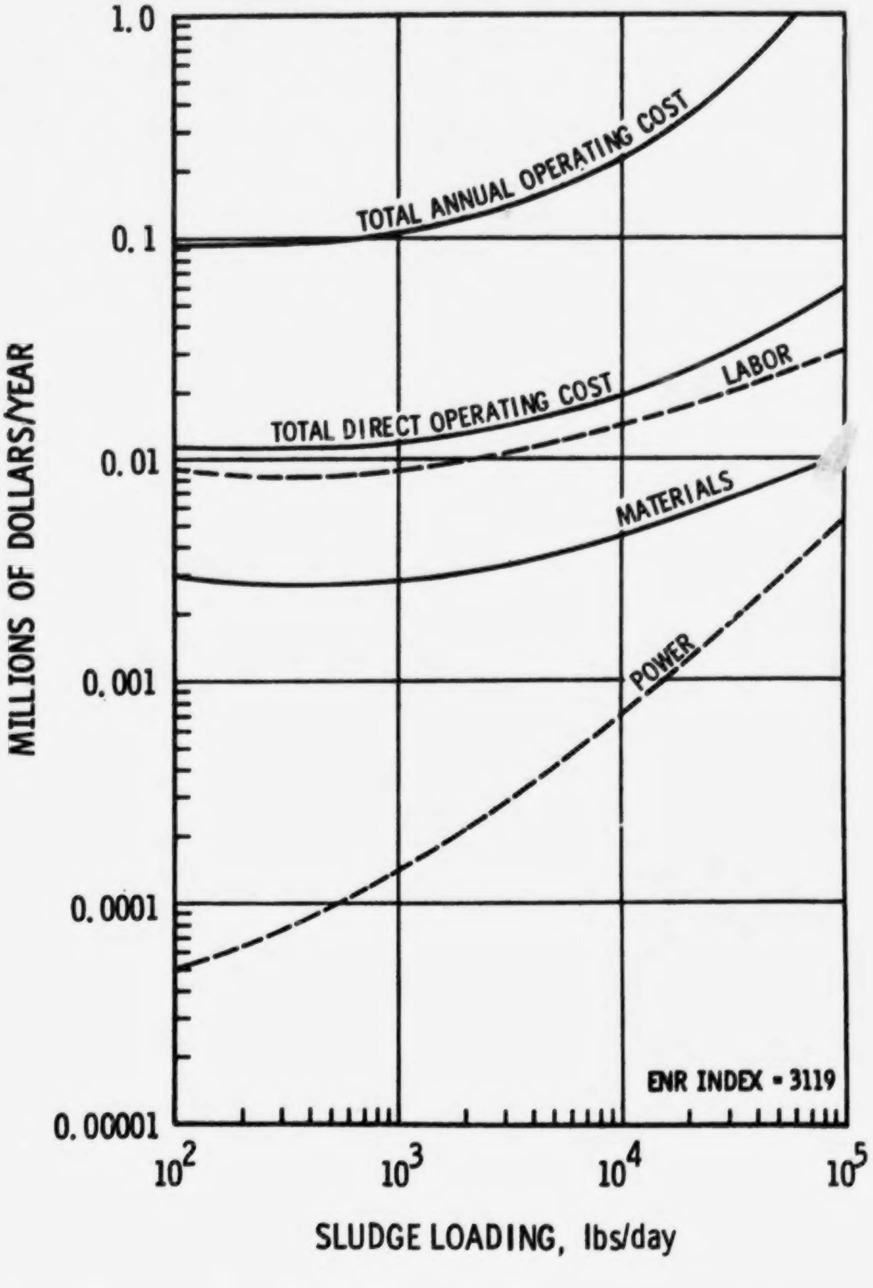
IV.6.5.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.6 CHEMICAL CONDITIONING [1]

IV.6.6.1 Description

Chemical conditioning is a process for coagulating sludge solids and releasing absorbed water.

The use of chemicals to condition sludge for dewatering is economical because of the increased yields and greater flexibility obtained.

Chemicals are most easily applied and metered in liquid form. Dissolving tanks are needed if the chemicals are received as dry powder. These tanks should be large enough for at least one-day's supply of chemicals and should be furnished in duplicate. They must be fabricated or lined with corrosion-resistant material. Polyvinyl chloride, polyethylene, and rubber are suitable materials for tank and pipe linings for handling acid solutions. Metering pumps, which must be corrosion resistant, are generally of the positive-displacement type with variable-speed or variable-stroke drives to control the flowrate. Another metering system consists of a constant-head tank supplied by a centrifugal pump. A rotameter and throttling valve are used to meter the flow.

The chemical dosage required for any sludge is determined in the laboratory. Filter-leaf test kits are used to determine chemical doses, filter yields, and the suitability of various filtering media. These kits have several advantages over the Büchner funnel procedure. In general, it has been observed that the type and concentration of sludge has the greatest impact on the quantity of chemical required. Sludges at less than 4 to 5% solids and difficult-to-dewater sludges require larger doses of chemicals and generally do not yield as dry a cake. Sludge types, listed in the approximate order of increasing chemical requirements for conditioning, are as follows:

Untreated (raw) primary sludge
Untreated mixed primary and trickling-filter sludge
Untreated mixed primary and waste activated sludge
Anaerobically digested primary sludge
Anaerobically digested mixed primary and waste activated sludge
Aerobically digested sludge (normally dewatered on drying beds without the use of chemicals for conditioning).

Intimate admixing of sludge and coagulant is essential for proper conditioning. The mixing must not break the floc after it has formed, and the detention is kept to a minimum so that sludge reaches the filter as soon after conditioning as possible. Mixing tanks are generally of the vertical type for small plants and of the horizontal type for large plants. They are ordinarily built of welded steel and lined with rubber or other acid-proof coating.

A typical layout for a mixing or conditioning tank has a horizontal agitator driven by a variable-speed motor to provide a shaft speed of 4 to 10 r/min. Overflow from the tank is adjustable to vary the detention period. Vertical cylindrical tanks with propeller mixers are also used.

IV.6.6.2 Typical Design Criteria

The dosage of inorganic chemicals for various types of sludges for vacuum filtration is shown below (conditioners are shown in percentage of dry sludge). Due to ease of handling, both cationic and anionic polymers are finding increased usage in view of inorganic chemicals.

	Fresh solids		Digested		Elutriated, digested	
Type of sludge	$FeCl_3$	CaO	$FeCl_3$	CaO	$FeCl_3$	CaO
Primary	1-2	6-8	1.5-3.5	6-10	2-4	
Primary and trickling filter	2-3	6-8	1.5-3.5	6-10	2-4	
Primary and activated	1.5-2.5	7-9	1.5-4	6-12	2-4	
Activated (alone)	4-6					

IV.6.6.3 Costs

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes primary sludge pumps, clarifiers, liquid ferric chloride, chemical feed equipment sized for twice the average feed rate, storage for at least 15 days, rapid mix tank, stainless steel mixer, and building (except for plants with a capacity less than 1 Mgal/d). The rapid mix tank is constructed of concrete, and multiple basins are used for volumes greater than 1,500 ft³. Costs are based on primary circular clarification with ferric chloride addition. A ferric chloride dosage of 100 mg/L was assumed for cost estimation. A dosage of 5% of ferric chloride was used as a percentage of dry primary sludge in deriving the chemical cost.

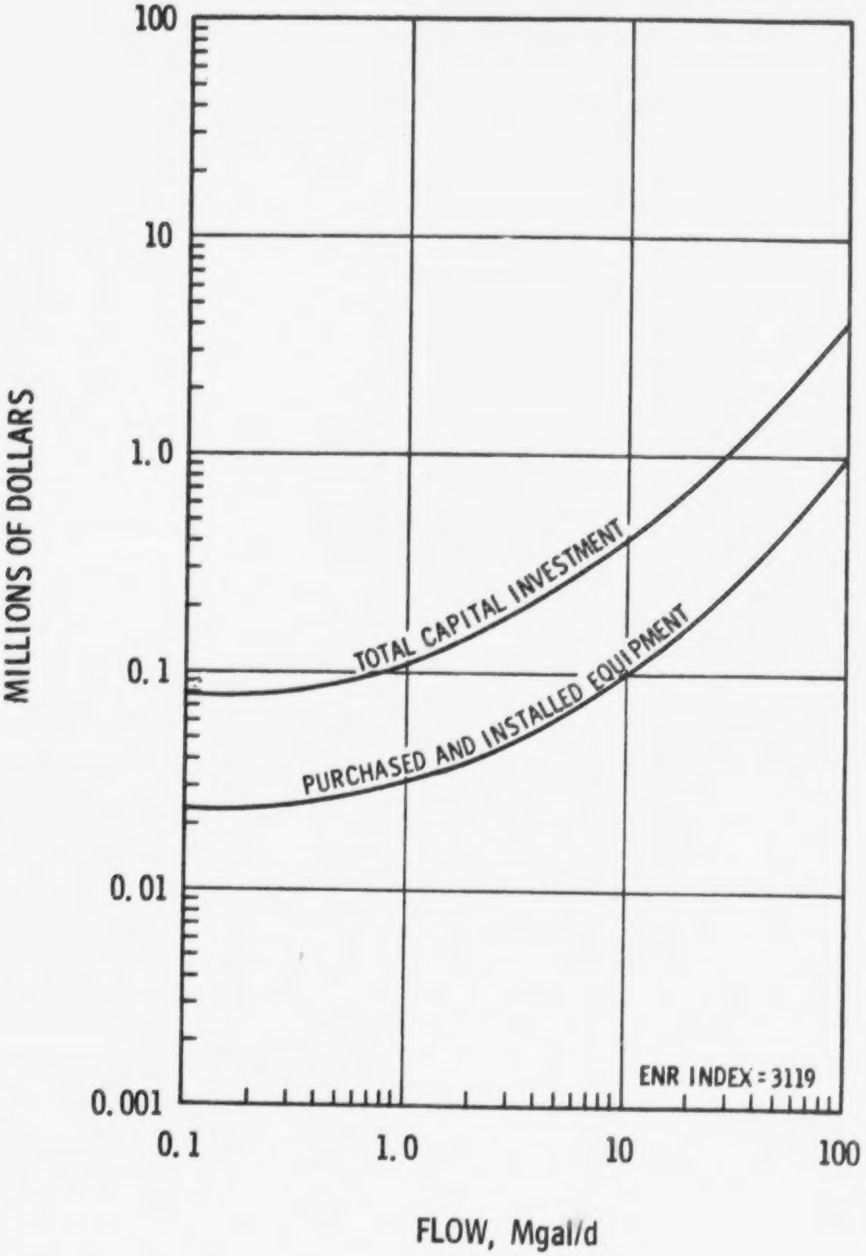
To adjust costs at the effective flow (Q_E) for different $FeCl_3$ dosages, use the following equation:

$$Q_E = Q_{Design} \times \frac{FeCl_3\text{dosage}}{100\ \text{mg/L}}$$

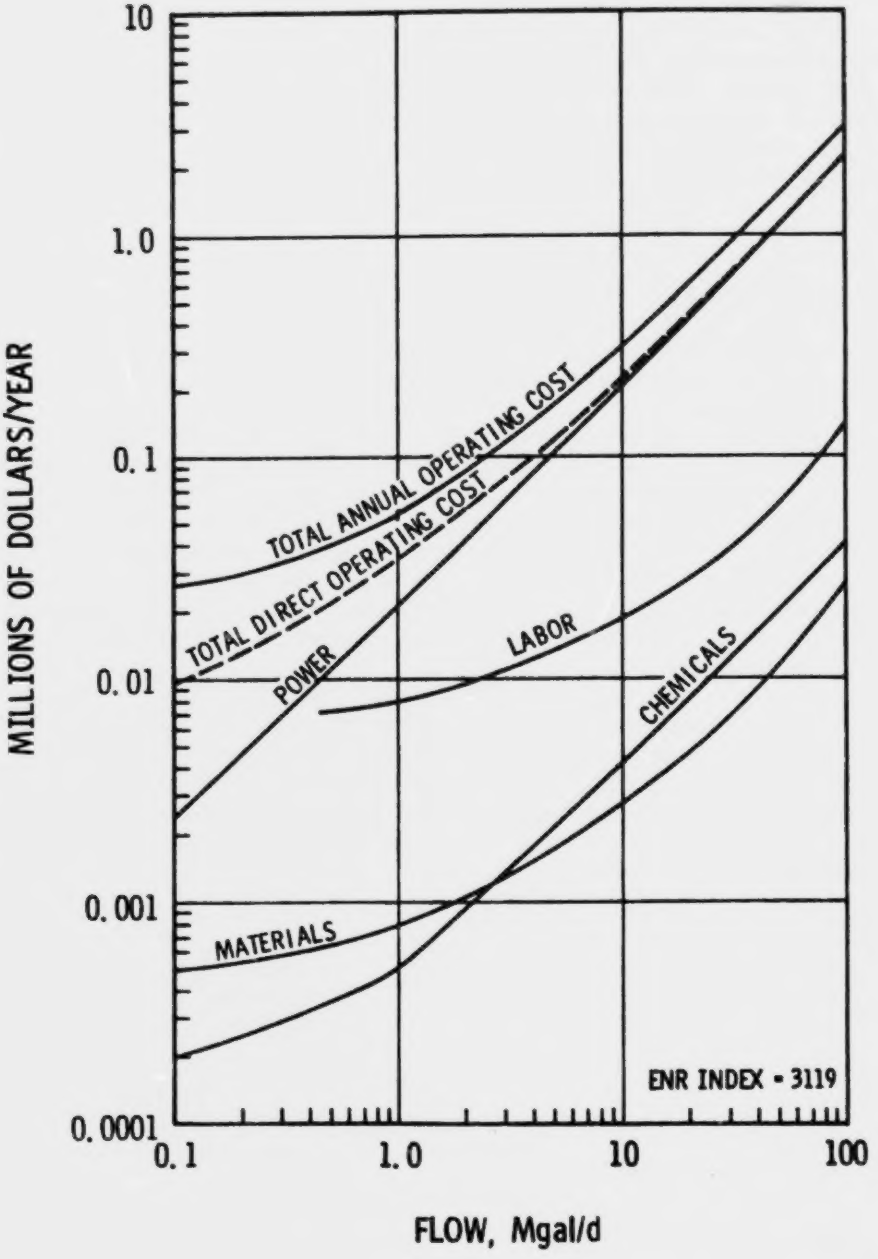
IV.6.6.4 References

1. Metcalf and Eddy, Wastewater Engineering - Treatment, Disposal, Reuse, McGraw-Hill, Inc., 1979. pp. 634-636.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.7 THERMAL CONDITIONING (HEAT TREATMENT OF SLUDGE) [1]

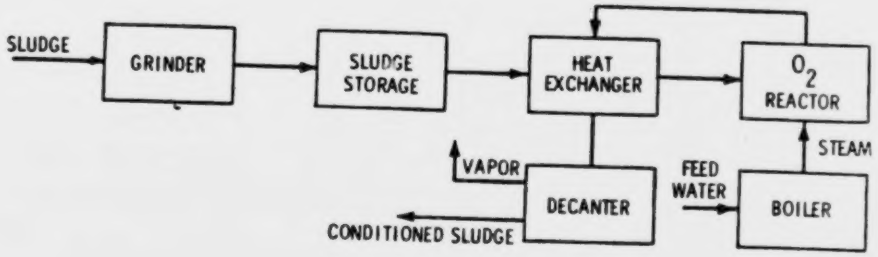
IV.6.7.1 Description

Heat treatment is essentially a conditioning process that prepares sludge for dewatering on vacuum filters or filter presses without the use of chemicals.

The heat treatment process involves heating sludge to 144°C to 210°C for short periods of time under pressure of 150 to 400 lb/in.^{2} gage. In addition, the sludge is sterilized and generally stabilized and rendered inoffensive. Heat treatment results in coagulation of solids, a breakdown in the cell structure of sludge, and a reduction of the water affinity of sludge solids.

Several proprietary variations exist for heat treatment. In these systems, sludge is passed through a heat exchanger into a reactor vessel, where steam is injected directly into the sludge to bring the temperature and pressure into the necessary ranges. In one variation, air is also injected into the reactor vessel with the sludge. The detention time in the reactor is approximately 30 minutes. After heat treatment, the sludge passes back through the heat exchanger to recover heat, and then is discharged to a thickener-decant tank. The thickened sludge may be dewatered by filtration or centrifugation to a solids content of 30 to 50 percent. The sludge may be ground prior to heat treatment. A small quantity (∿5% of that required for complete oxidation may be introduced into the reaction to minimize reduced sulfur type odor.

Complete heat treatment systems are generally proprietary, and the most common systems are supplied by five manufacturers. The major equipment common to these processes are grinders, sludge feed pumps and handling equipment, heat exchangers, reactors, boilers, and separators. A flow diagram of the heat treatment system is shown below.



(SEE SECTION IV.7)

IV.6.7.2 Typical Design Criteria

Typical design criteria for a heat treatment system are listed below.

Design criteria	Value
Temperature	140-210°C
Pressure	150-400 lb/in.^{2} gage
Detention time	30-90 min
Steam consumption	600 lb/1,000 gal of sludge

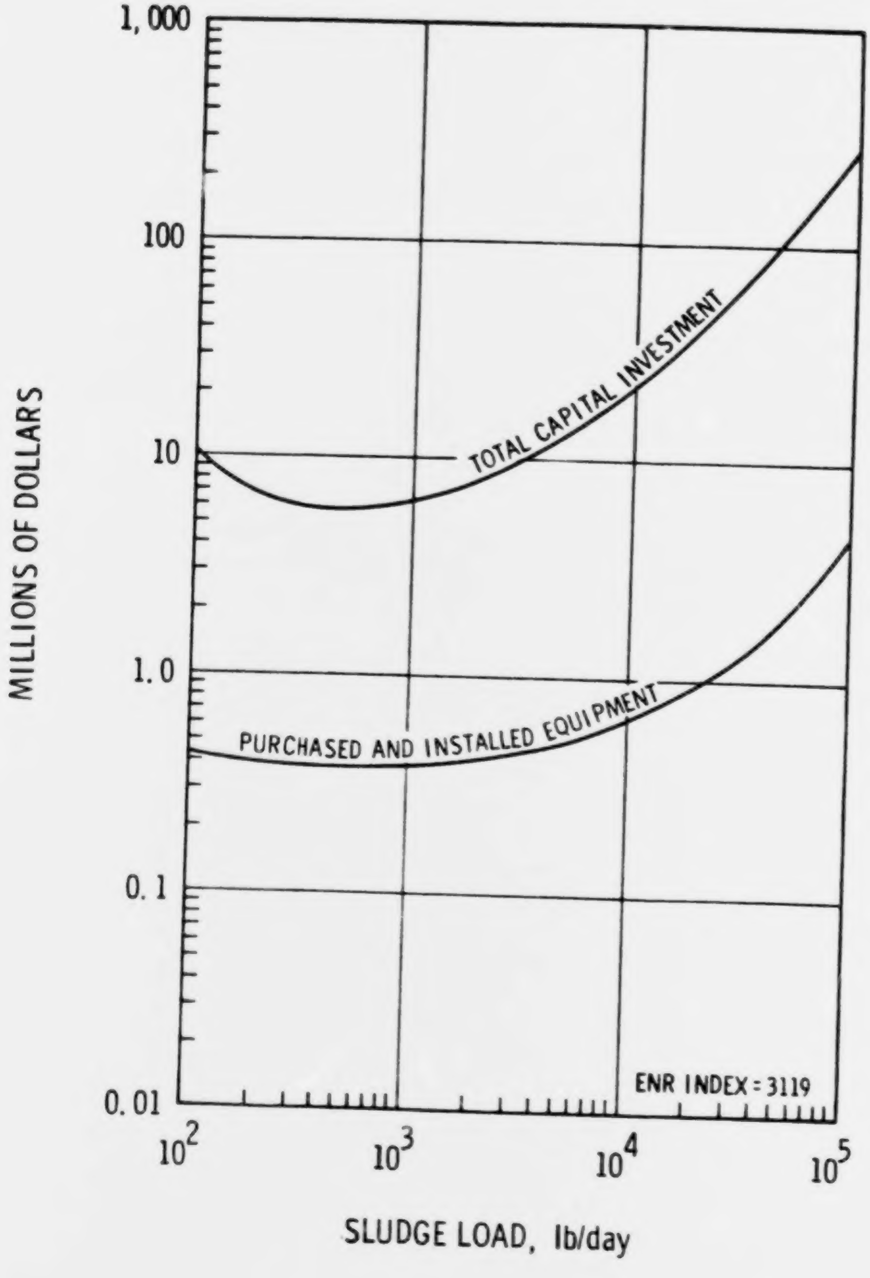
IV.6.7.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes sludge feed pumps, grinders, heat exchangers, reactors, boilers, gas separators, and buildings. Fuel costs are for steam generation. Sludge is assumed to be 4.5% solids with the system operating 3,000 hr/yr.

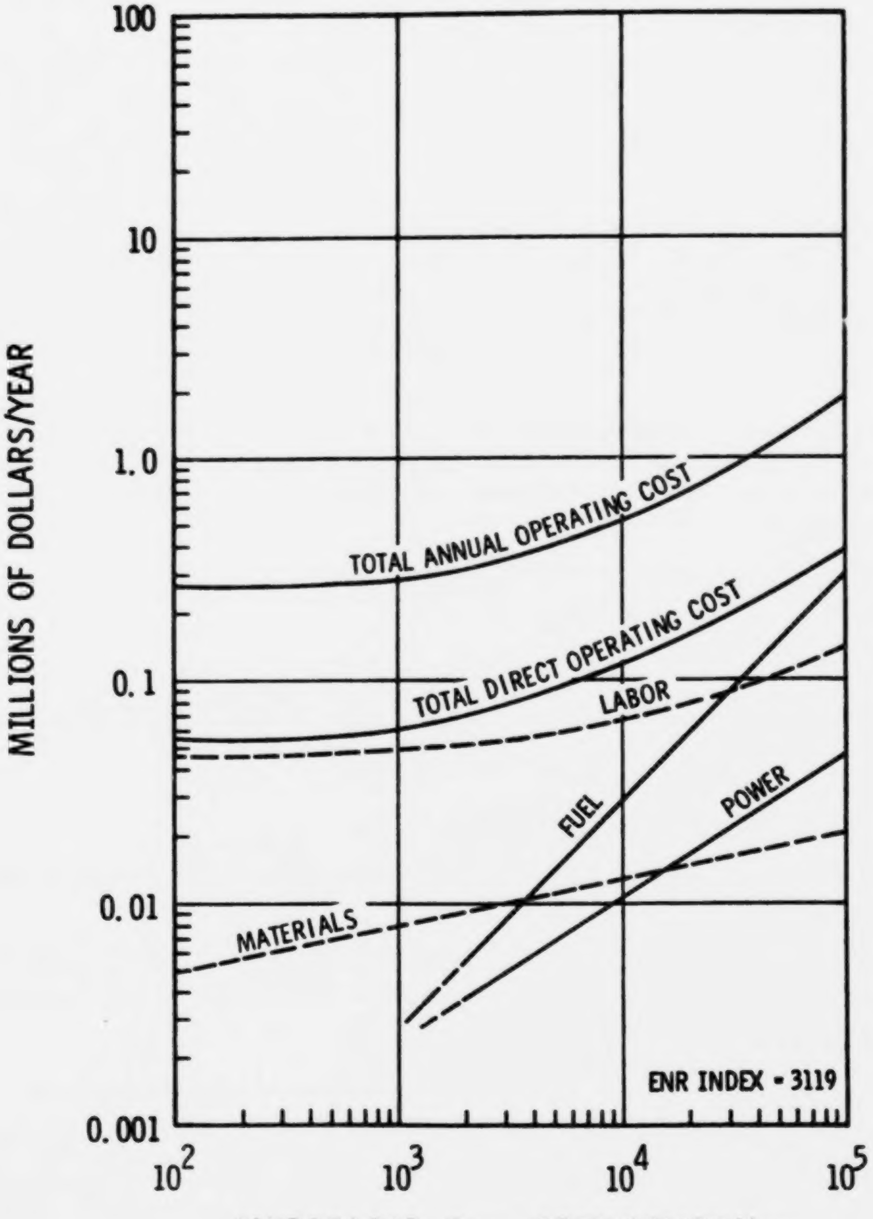
IV.6.7.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



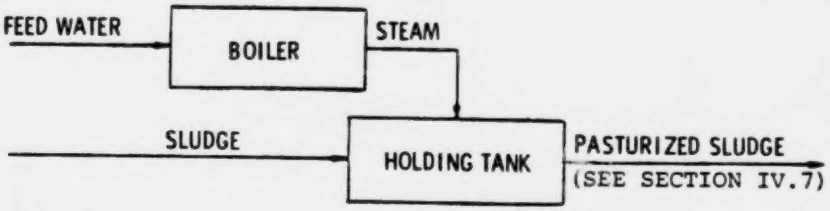
IV.6.8 DISINFECTION (HEAT) [1]

IV.6.8.1 Description

Heating to pasteurization temperatures is a well known method of destroying pathogenic organisms that has been applied successfully for disinfecting sludge. Pasteurization implies heating to a specific temperature for a time period sufficient to destroy undesirable organisms in sludge and to make sludge suitable for land disposal on cropland. Usually heat is applied at 70 to 75°C for 20 to 60 minutes. Treatment can be applied to raw liquid sludge (thickened or unthickened), or stabilized or digested sludge.

Pasteurization is usually a batch process, consisting of a reactor to hold sludge, a heat source, and heat exchange equipment, pumping and piping, and instrumentation for automated operation. Pasteurization has little effect on sludge composition or structure because the sludge is only heated to a relatively moderate temperature.

Equipment normally associated with the heating process includes sludge handling and control equipment, heating equipment, and instrumentation. A flow diagram of the pasteurization system is shown below.



IV.6.8.2 Typical Design Criteria

Typical design criteria for a pasteurization process are shown below.

Design criteria	Value
Temperature	70-75°C
Time	20-60 min
Heat required	4-6 x 10^6 Btu/ton of sludge solids
Operation	Two units or more are usually designed in parallel so that one unit can be filling while the other is holding sludge for the required time; units can share a common boiler.

IV.6.8.3 Costs

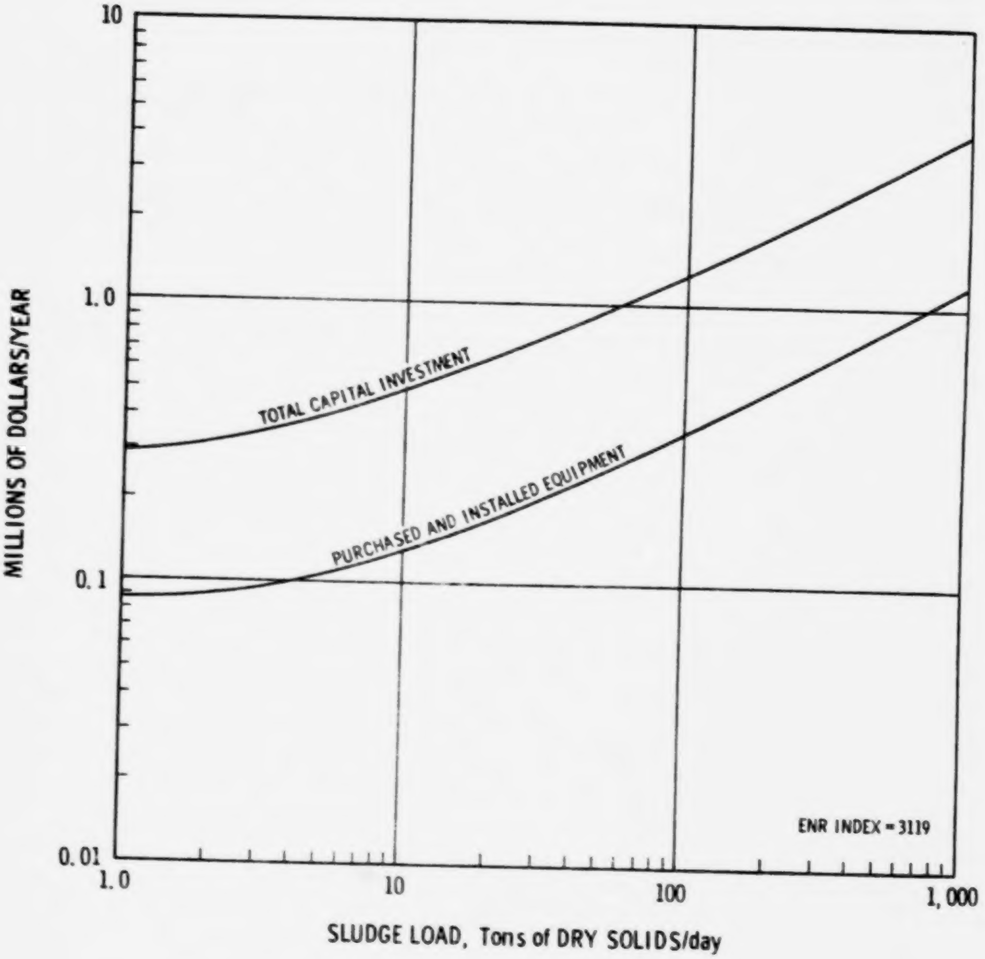
Purchased equipment and installation cost for estimation of the total capital investment includes steam boiler, pasteurization tanks, sludge pumping, and automatic controls. The following design characteristics were assumed for cost estimation:

Design characteristic	Assumed value
Sludge temperature	17°C
Sludge solids	5%
Pasteurization temperature	70°C
Pasteurization time	1 hr
Capacity	Single tank up to 10 tons of sludge solids per day; two tanks above.

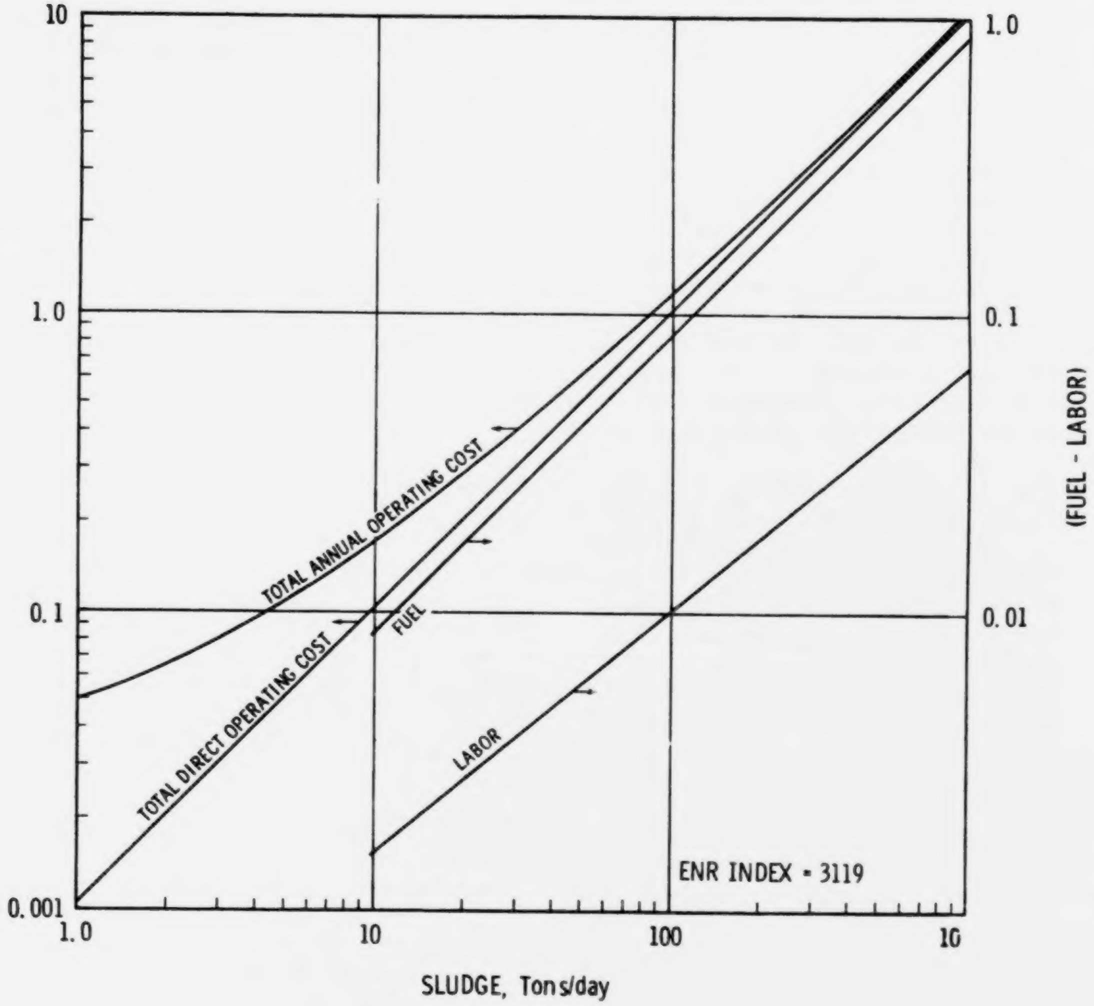
IV.6.8.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.9 VACUUM FILTRATION [1]

IV.6.9.1 Description

Vacuum filters are used to dewater sludges so as to produce a cake having the physical handling characteristics and moisture contents required for subsequent processing.

A rotary vacuum filter consists of a cylindrical drum rotating partially submerged in a vat or pan of conditioned sludge. The drum is divided radially into a number of sections, which are connected through internal piping to ports in a valve body (plate) at the hub. This plate rotates in contact with a fixed valve plate with similar ports, which are connected to a vacuum supply, a compressed air supply, and an atmospheric vent. As the drum rotates, each section is thus connected to the appropriate service. Various operating zones are encountered during a complete revolution of the drum. In the pickup or form section, vacuum is applied to draw liquid through the filter covering (media) and form a cake of partially dewatered sludge. As the drum rotates, the cake emerges from the liquid sludge pool, while suction is maintained to promote further dewatering. A lower level of vacuum often exists in the cake drying zone. A scraper blade may be provided for sludge removal, followed by a wastewater spray to clean the media.

The three principal types of rotary vacuum filters are the drum type, coil type, and the belt type. The filters differ primarily in the type of covering used and the cake discharge mechanism employed. Cloth media are used on drum and belt types; stainless steel springs are used on the coil type. Infrequently, a metal media is used on belt types. The drum filter also differs from the other two in that the cloth covering does not leave the drum but is washed in place, when necessary. The design of the drum filter provides considerable latitude in the amount of cycle time devoted to cake formation, washing, and dewatering; the design also minimizes inactive time.

The top feed drum filter is a variation of the conventional drum filter. In this case, sludge is fed to the vacuum filter through a hopper located above the filter. The potential advantages of the top feed drum filter are that gravity aids in cake formation; capital costs may be lower since the feed hopper is smaller and no sludge agitator and related drive equipment are required; and "blinding" of the media may be reduced.

The coil-type vacuum filter uses two layers of stainless steel coils arranged in corduroy fashion around the drum. After a dewatering cycle, the two layers of springs leave the drum and are separated from each other so that the cake is lifted off the lower layer of springs and discharged from the upper layer. Cake release is essentially free of problems. The coils are then washed and reapplied to the drum. The coil filter has been and is widely used for all types of sludge. However, sludges with particles that are both extremely fine and resistant to flocculation dewater poorly on coil filters.

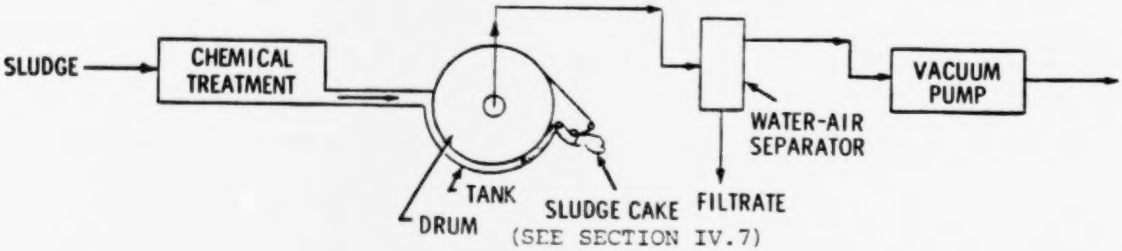
Solids loss during cake formation and during media washing produces high filtrate suspended solids for return to liquid treatment processes.

Media on the belt-type filter leaves the drum surface at the end of the drying zone and passes over a small diameter discharge roll to facilitate cake discharge. Washing of the media next occurs before it returns to the drum and to the vat for another cycle. This type filter normally has a small diameter curved bar between the point where the belt leaves the drum and the discharge roll that aids in maintaining belt dimensional stability. In practice, it is frequently used to insure adequate cake discharge.

Many types of filter media are available for belt and drum filters. There is some question whether increases in yield due to operating vacuums greater than 15 inches of mercury are justifiable. The cost of a greater filter area must be balanced against the higher power costs for higher vacuums. An increase from 15 to 20 inches of vacuum is reported to have provided about 10 percent greater yield in three full-scale installations.

Chemical conditioning is often employed to agglomerate a large number of small particles. It is almost universally applied with mixed sludges.

Equipment normally associated with the vacuum filtration process includes the rotary vacuum filter, vacuum pump, filtrate receiver, filtrate pump, sludge conditioning apparatus, and sludge conveyors. A flow diagram for a vacuum filtration system is shown below.



(SEE SECTION IV.7)

IV.6.9.2 Typical Design Criteria

Typical design criteria for vacuum filtration are shown below.

Design criterion	Value
Typical loading for raw primary sludges	7-15 lb dry solids/hr/ft^2
Typical loading for digested primary sludges	4-7 lb dry solids/hr/ft^2
Typical loading for mixed digested sludges	3.5-5 lb dry solids/hr/ft^2

Loading is a function of feed solids concentrations, subsequent processing requirements, and chemical preconditioning.

IV.6.9.3 Costs

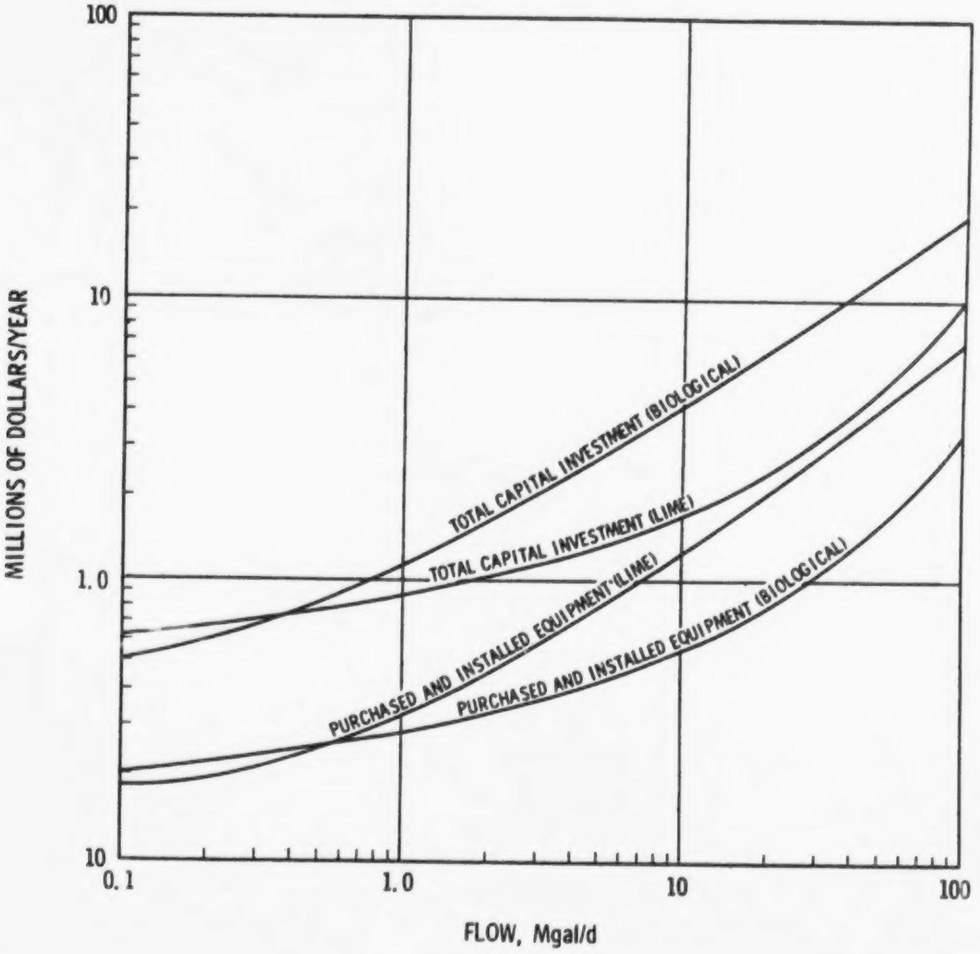
Purchased equipment and installation cost for estimation of the total capital investment includes pumps, internal piping and electrical controls, mechanical equipment, conveyors, sludge cake storage hopper, building, and chemical handling and storage facilities. Costs are for dewatering of combined primary and secondary digested sludge consisting of 900 lb dry solids/Mgal plant flow or lime sludge consisting of 4,500 lb dry solids/Mgal plant flow. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Filter yield	5 lb/ft^2 for biological sludge; 8 lb/ft^2 for lime sludge
Operation time (excluding down time for maintenance)	6 hr/d for 1 Mgpd or less; 12 hr/d for 10 Mgpd
Chemical dosage	
$FeCl_3$	35 lb/Mgal
CaO	90 lb/Mgal

IV.6.9.4 References

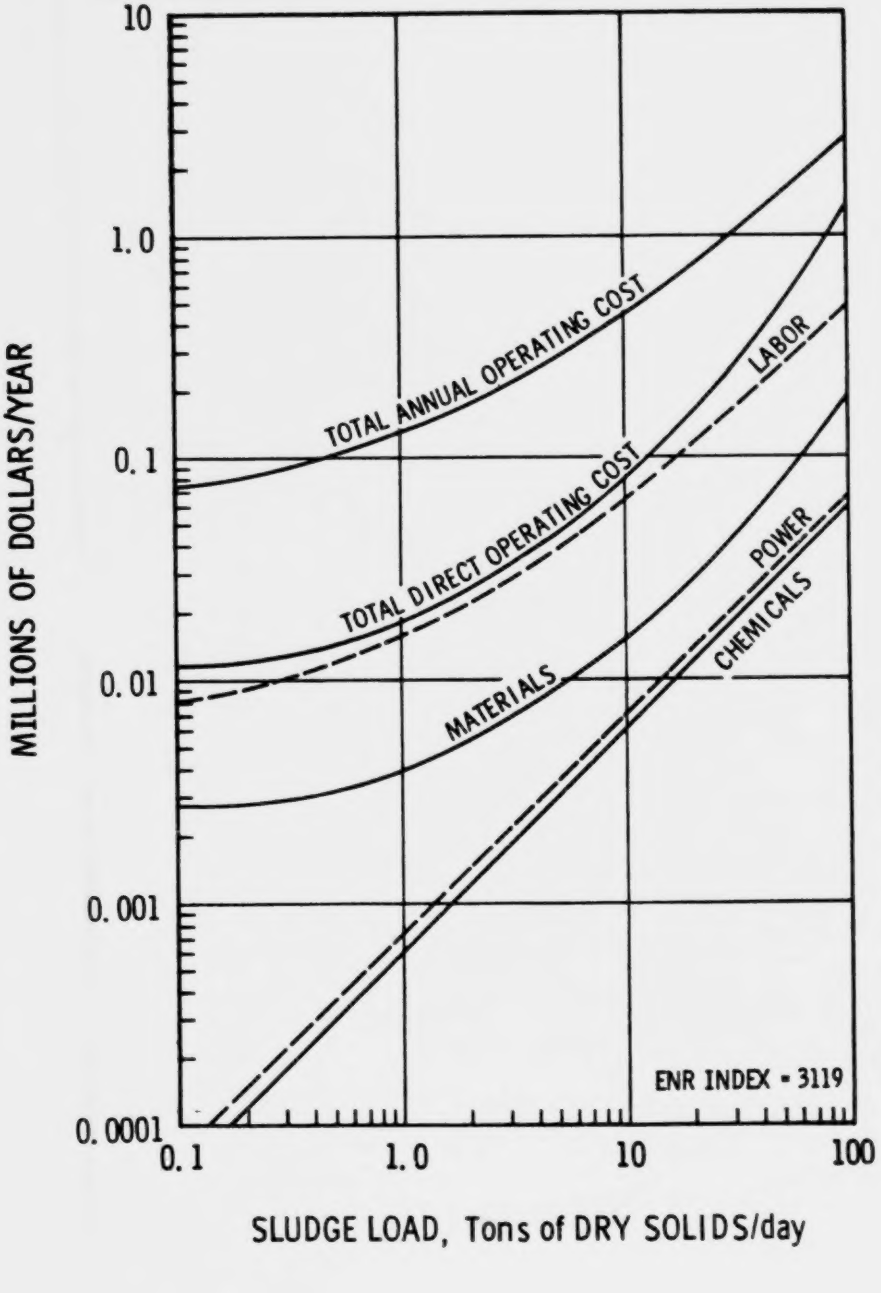
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



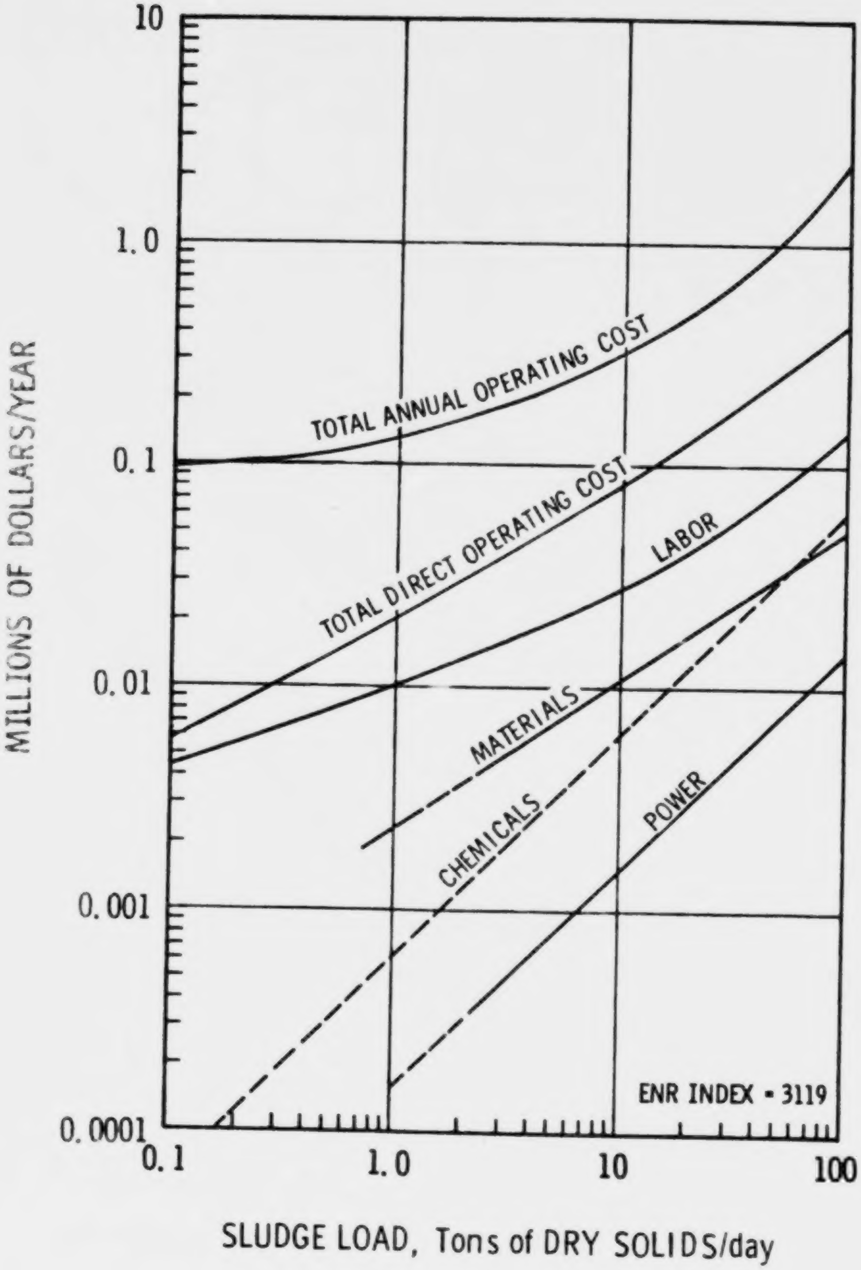
ANNUAL OPERATING COST

(LIME SLUDGE)



ANNUAL OPERATING COST

(BIOLOGICAL SLUDGE)



IV.6.10 FILTER PRESS DEWATERING [1]

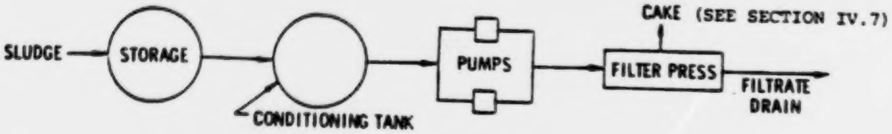
IV.6.10.1 Description

The diaphragm filter press is a recent extension of filter press technology to increase the throughput of a press and provide a higher solids content in wastewater filter cake. The press makes use of a rubber diaphragm with pressurized water to provide a high pressure on the partially dewatered sludge cake in the press after conventional dewatering methods have been used. This diaphragm provides the support for the filter cloth on one side of the press cavity.

Filtration of the sludge is accomplished by charging the chambers of the press with sludge under pump pressure in the conventional manner, but at a generally lower pressure, and allowing a cake to develop. The time alloted to this cycle is dependent upon the characteristics of the sludge, but is scheduled to continue only as long as high filtration rates are in progress. This cycle usually is in the 10- to 30-minute range. The pump pressurizing system is then turned off, and water pressure is applied inside the diaphragm. This pressure, in the 200-psi range, applies a uniform pressure over the cake and further reduces the water content. The squeezing cycle has been shown to substantially reduce the overall cycle time for the press yet produce a low-moisture-content cake. The filter cake produced is thinner than that obtained in the conventional press but has a uniform moisture content in contrast to the conventional press. The reduction in operating time for the sludge pumps is reported to substantially reduce wear and required maintenance. The pressurized water for the diaphragm actuation is a closed recycle system; therefore, components operate under predictable conditions and no effluent is produced. Diaphragm presses are designed to make use of a number of automation features to reduce labor and recycle time.

Common modifications include opening of the filter press to allow simultaneous discharge of filter cake from all cavities, rejection of the sludge cake by physical movement of the filter cloth from vibration or actual movement of the cloth in a forced rejection mode by partial withdrawal of the filter cloth loop, automatic washing of filter cloths at each cycle or as conditions dictate, air purging of feed and filtrate lines between cycles, and full automation of press operation.

Equipment normally associated with the filter press dewatering process includes diaphragm filter presses, sludge pumps, cake conveyors, and sludge conditioning tanks. A flow diagram of a filter press system is shown on the following page.



IV.6.10.2 Typical Design Criteria

Typical design criteria for a diaphragm filter press are shown below.

Design criterion	Value
Filter areas	5-40 ft^2/chamber
Cake thickness	1/2-3/4 in.
Sludge yield	0.4-0.8 lb/ft^2 of filter area
Total cycle time	20-50 min

IV.6.10.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes the chemical feed system, sludge feed pump, dewatering unit with all necessary accessories, and a conveyor system to transport cake to the next process.

A cost estimate has been derived for use of a diaphragm filter press in a large multiple-press installation with a sludge cake production of 250 dry tons/d. Sludge was assumed to be a mix of 2:1 secondary to primary with a feed solids of 5%. Sludge conditioning was assumed to be lime at 20% and $FeCl_3$ at 7% dry sludge solids. Lime cost was assumed at $44/ton and $FeCl_3$ at $130/ton. Pricing is based on the largest size presses available. The total number includes one spare. The capital cost (1978 dollars) includes the chemical feed system, sludge feed pumps, dewatering unit with all necessary accessories, and a conveyor system to transport cake to the next process. The total installation cost was obtained by utilizing a multiplication factor of 3 which includes installation, piping, utilities, building, and engineering. Labor is assumed at $21,000/manyear.

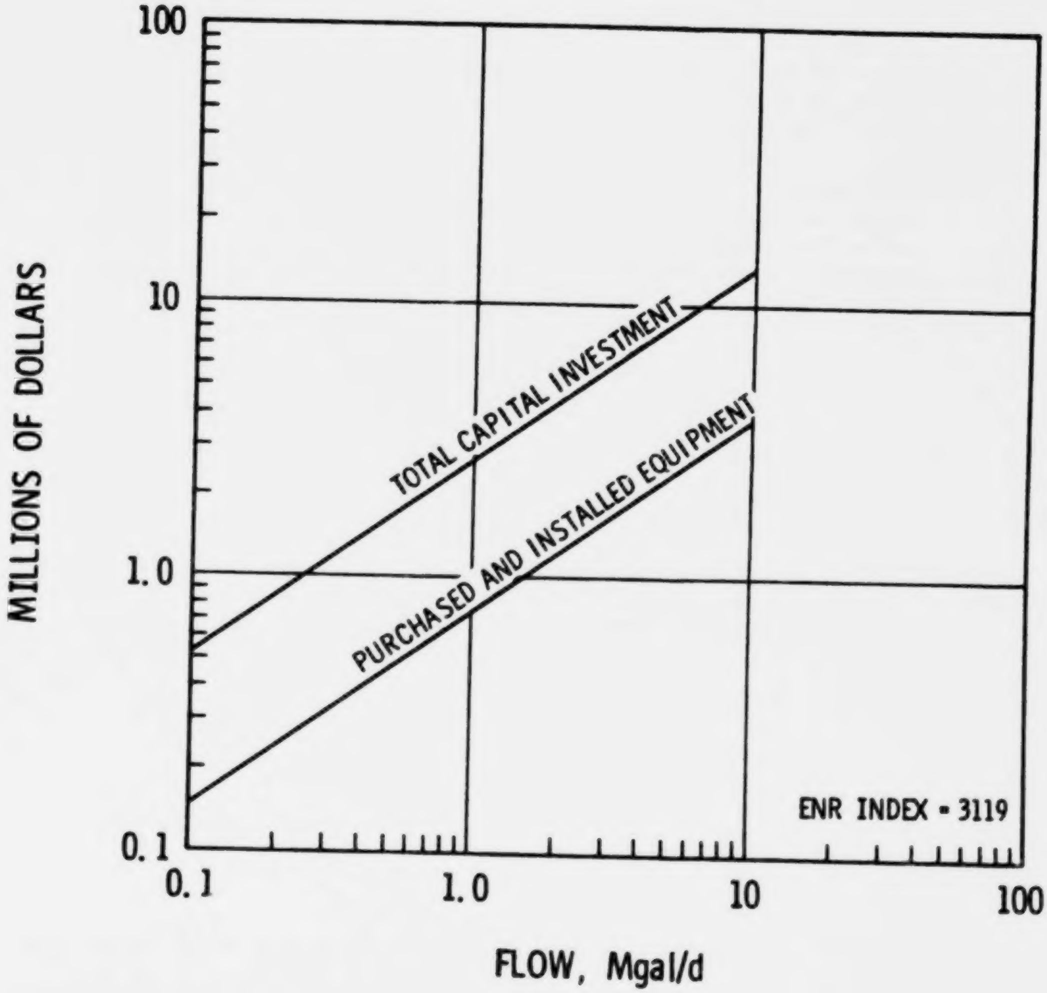
Item	Cost
Cost summary[a]	
Capital cost	$19,500,000
Lime system	1,000,000
$FeCl_3$ system	500,000
Conveyors	2,000,000
Total	$23,000,000
Annual costs[a]	
Amortization at 9%	$ 2,070,000
Chemicals	1,633,000
Power at 66.7 kWh/ton and $0.04/kWh	243,000
Water 12×10^6 gallons	6,400
Labor - operation	504,000
Maintenance -	
Cloth and diaphragm replacement-materials	312,000
Labor replacement	31,500
Equipment maintenance at 2% of purchase cost	153,000
Total annual costs	$ 4,953,000
Unit cost/dry ton of sludge cake	$ 54.28

[a] Costs are in 1978 dollars.

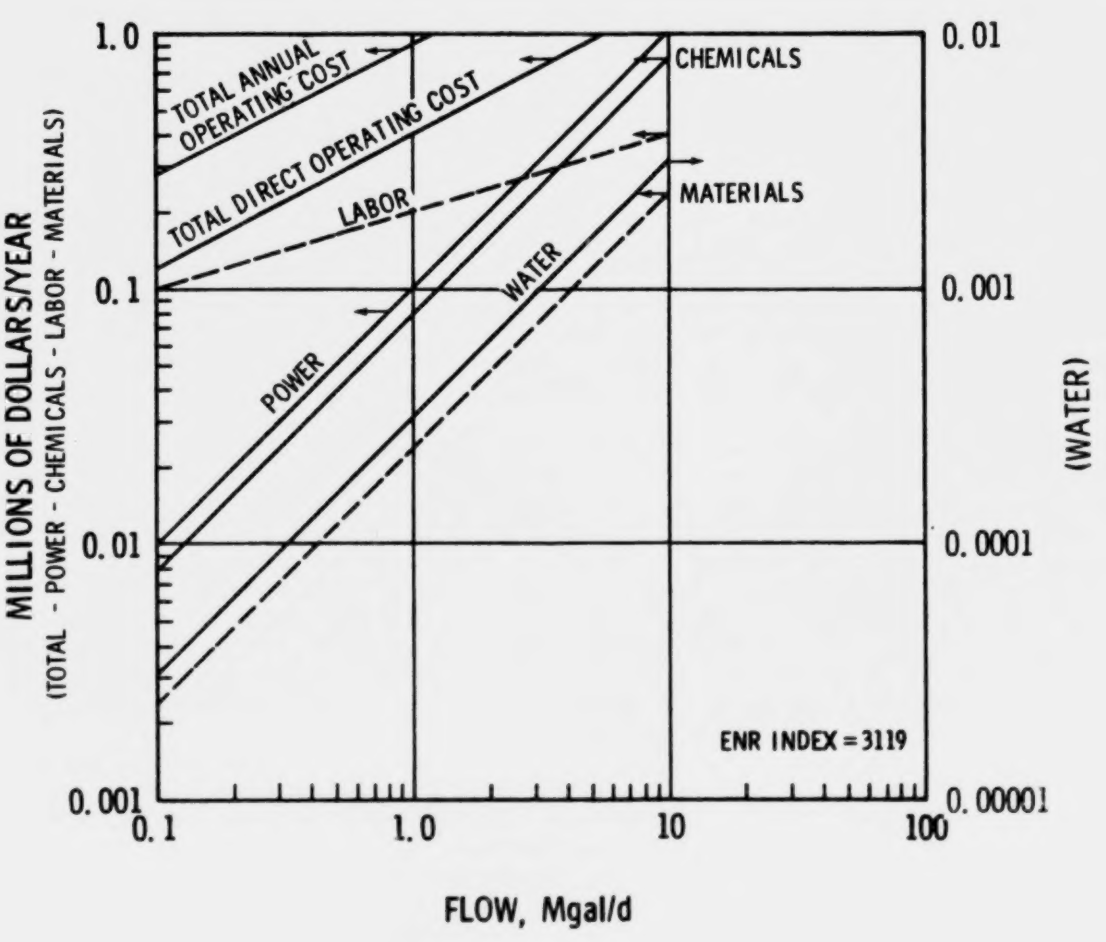
IV.6.10.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.11 BELT FILTER DEWATERING [1]

IV.6.11.1 Description

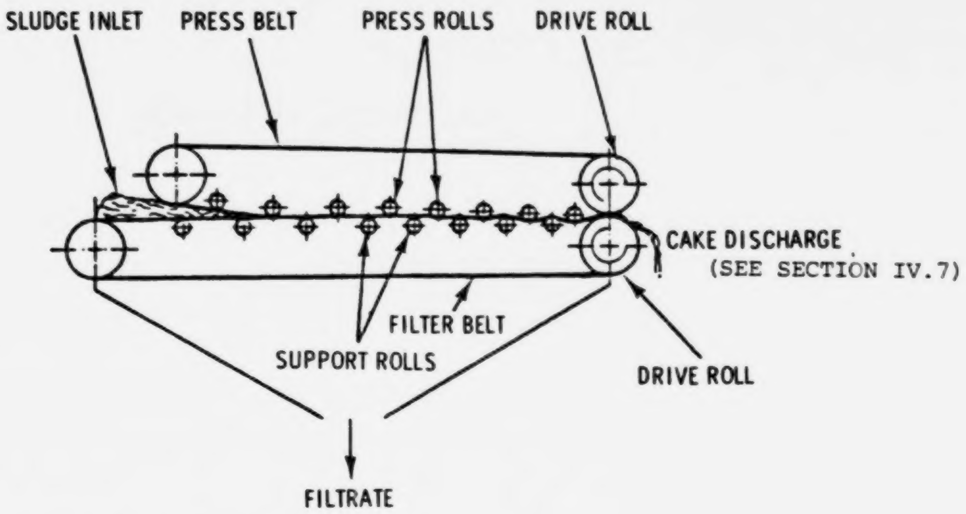
Belt filter dewatering is the removal of water from sludge using filtration in the form of rolling belts.

A belt filter consists of an endless filter belt that runs over a drive and guide roller at each end like a conveyor belt. The upper side of the filter belt is supported by several rollers. Above the filter belt is a press belt that runs in the same direction and at the same speed; its drive roller is coupled with the drive roller of the filter belt. The press belt can be pressed on the filter belt by means of a pressure roller system whose rollers can be individually adjusted horizontally and vertically. The sludge to be dewatered is fed on the upper face of the filter belt and is continuously dewatered between the filter and press belts. After having passed the pressure zone, further dewatering in a reasonable time cannot be achieved by only applying static pressures; however, a superimposition of shear forces can effect this further dewatering. The supporting rollers of the filter belt and the pressure rollers of the pressure belt are adjusted in such a way that the belts and the sludge between them describe an S-shaped curve. Thus, there is a parallel displacement of the belt relative to each other due to the differences in the radii. After further dewatering in the shear zone, the sludge is removed by a scraper.

Some units consist of two states; the initial draining zone is on the top level, followed by an additional lower section wherein pressing and shearing occur. A significant feature of the belt filter press is that it employs a coarse-mesh, relatively open weave, metal-medium fabric. This is feasible because of the rapid and complete cake formation obtainable when proper flocculation is achieved. Belt filters do not need vacuum systems and do not have the sludge pickup problem occasionally experienced with rotary vacuum filters. The belt filter press system includes auxiliaries such as polymer solution preparation equipment and automatic process controls.

Some belt filters include the added feature of vacuum boxes in the free drainage zone. To obtain higher cake solids, a vacuum of about 6 in. Hg is applied. A "second generation" of belt filters has extended shearing or pressure states that produce substantial increases in cake solids but are more costly.

Equipment normally associated with belt filter dewatering processes includes the belt filter, chemical feed equipment, cake conveyors, and sludge pump. A flow diagram of a belt filter is shown on the following page.


SLUDGE INLET
PRESS BELT
PRESS ROLLS
DRIVE ROLL
CAKE DISCHARGE
(SEE SECTION IV.7)
FILTER BELT
SUPPORT ROLLS
DRIVE ROLL
FILTRATE

IV.6.11.2 Typical Design Criteria

The loadings shown below are based on active belt area:

Sludge type	Sludge loading, gal/ft²/hr	Dry solids loading, lb/ft²/hr
Raw primary	27-34	13.5-17
Digested primary	20-24	20.5-24
Digested mixed/secondary	13-17	6.7-8.4

IV.6.11.3 Costs

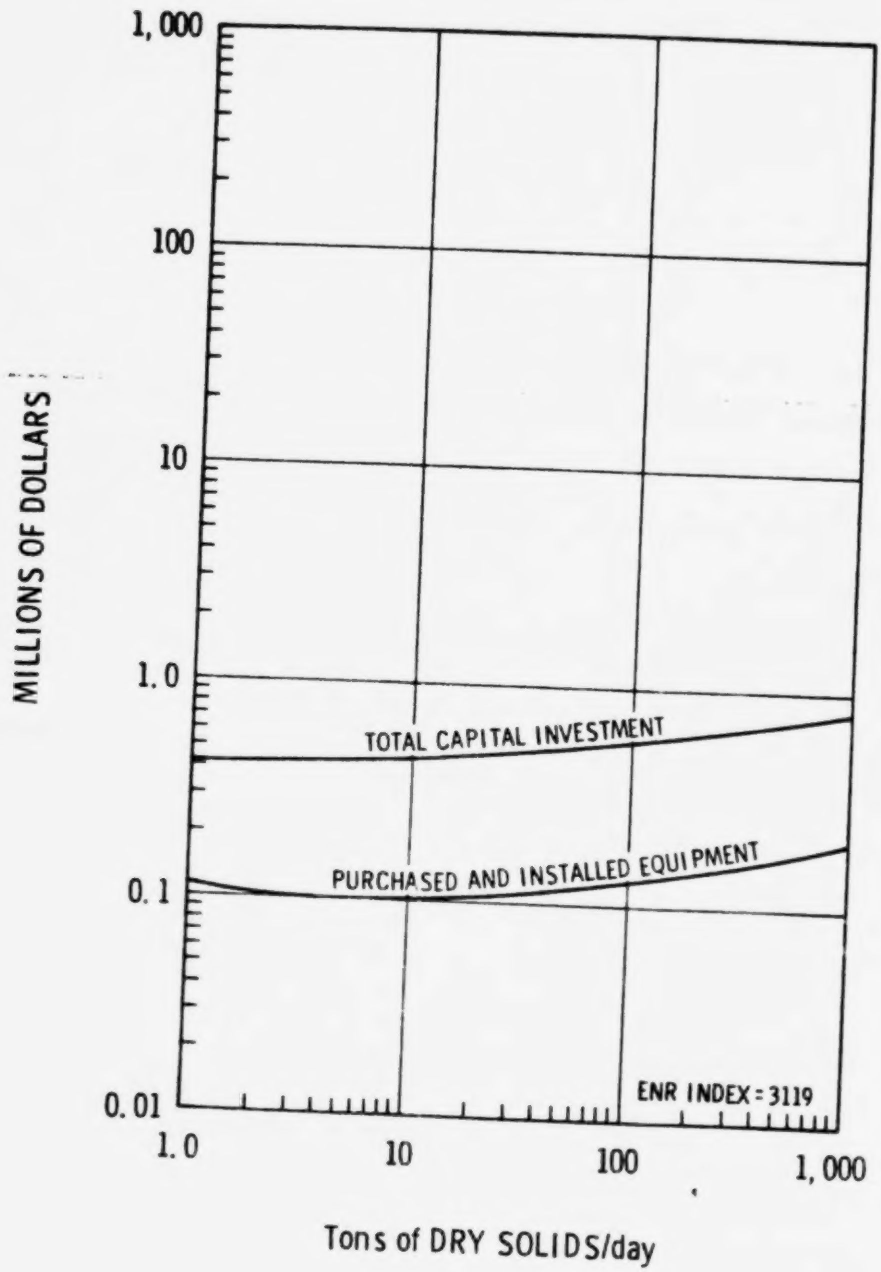
Purchased equipment and installation cost for estimation of total capital investment includes the belt filter press, sludge feed pumps, polymer pumps, and control panels. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Type of sludge	Primary and secondary (anaerobically digested) 5% solids concentration
Sludge production	7 d/wk
Dewatering operation	7 d/wk; 16 hr/d
Chemicals polymer	$17,520/yr at 8 tons/day

IV.6.11.4 References

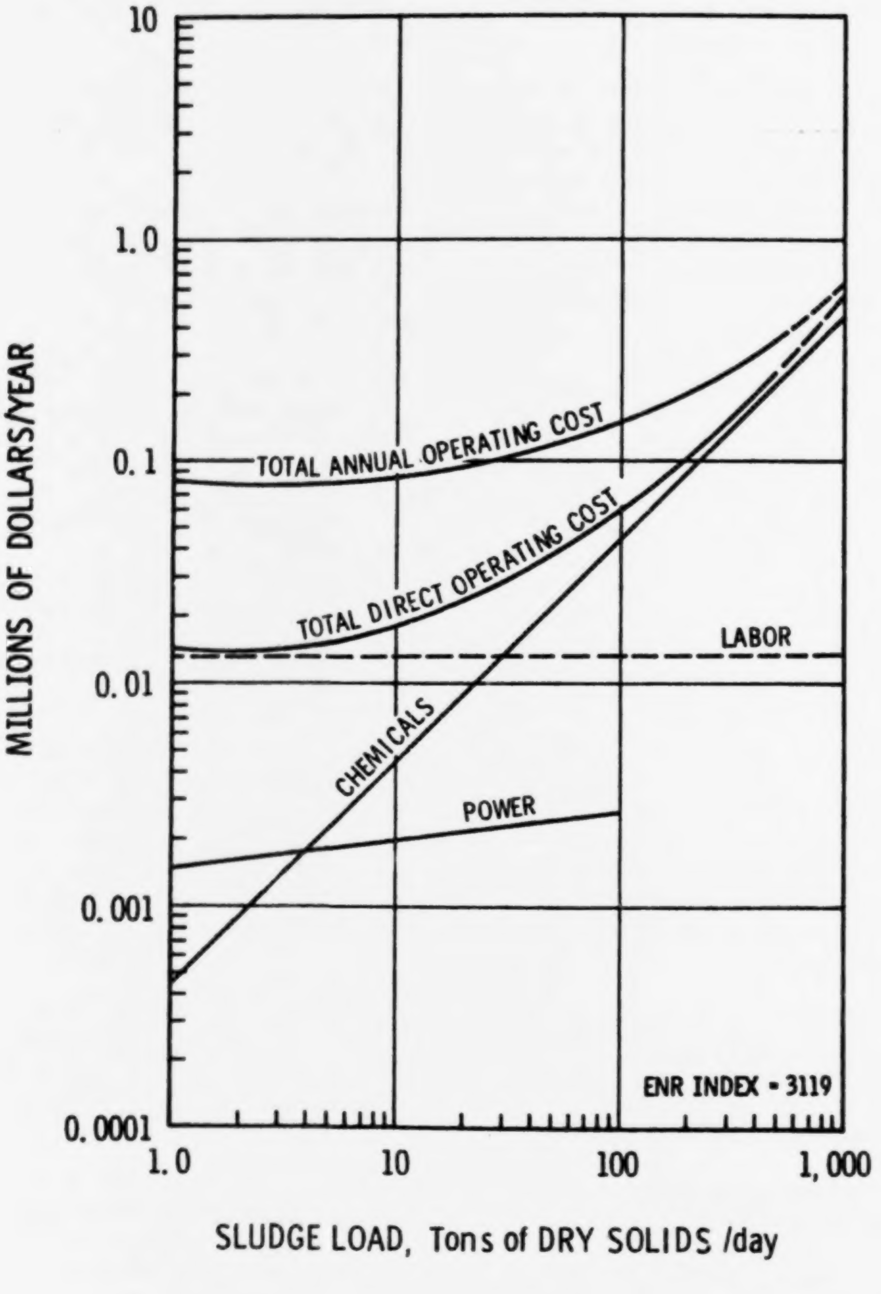
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



IV.6.12 CENTRIFUGAL DEWATERING [1]

IV.6.12.1 Description

Centrifuges are used to dewater sludges using centrifugal force to increase the sedimentation rate of sludge solids. The solid bowl, the disc, and the basket are the three most common types of units.

The solid-bowl continuous centrifuge assembly consists of a bowl and conveyor joined through a planetary gear system, designed to rotate the bowl and the conveyor at slightly different speeds. The solid cylindrical bowl, or shell, is supported between two sets of bearings and includes a conical section at one end. This section forms the dewatering beach over which the helical conveyor screw pushes the sludge solids to outlet ports and then to a sludge cake discharge hopper. The opposite end of the bowl is fitted with an adjustable outlet weir plate to regulate the level of the sludge pool in the bowl. The centrate flows through outlet ports either by gravity or by a centrate pump attached to the shaft at one end of the bowl. Sludge slurry enters the unit through a stationary feed pipe extending into the hollow shaft of the rotating bowl and passes to a baffled, abrasion-protected chamber for acceleration before discharge through the feed ports in the rotating conveyor hub into the sludge pool. Due to the centrifugal forces, the sludge pool takes the form of a concentric annular ring on the inside of the bowl. Solids settle through this ring to the wall of the bowl where they are picked up by the conveyor scroll. Separate motor sheaves or a variable speed drive can be used to adjust the bowl speed for optimum performance.

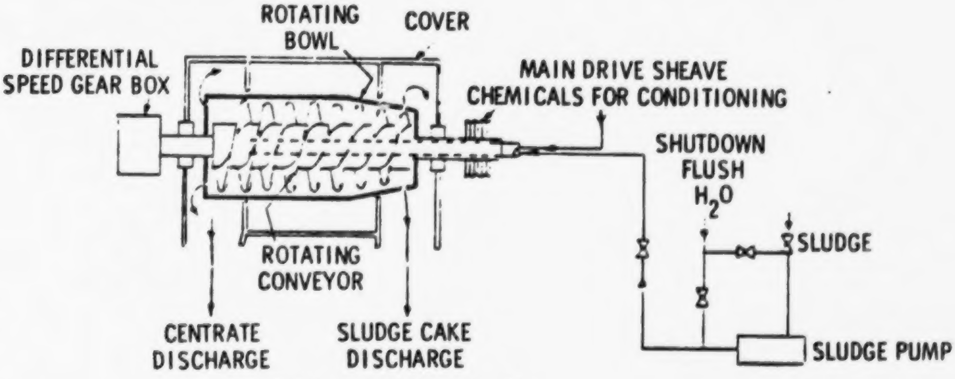
Bowls and conveyors can be constructed from a large variety of metals and alloys to suit special application. For dewatering of wastewater sludges, mild steel or stainless steel has been used normally. Because of the abrasive nature of many sludges, hardfacing materials are applied to the leading edges and tips of the conveyor blades, the discharge ports, and other wearing surfaces. Such wearing surfaces may be replaced by welding when required.

In the continuous concurrent solid-bowl centrifuge, incoming sludge is carried by the feed pipe to the end of the bowl opposite the discharge. Centrate is skimmed off and cake proceeds up the beach for removal. As a result, settled solids are not disturbed by incoming feed.

In the disc-type centrifuge, the incoming stream is distributed between a multitude of narrow channels formed by stacked conical discs. Suspended particles have only a short distance to settle so that small and low density particles are readily collected and discharged continuously through fairly small orifices in the bowl wall. The clarification capability and throughput range are high, but sludge concentration is limited by the necessity of discharging through orifices 0.050 in. to 0.100 in. in diameter. Therefore, it is generally considered a thickener rather than a dewatering device.

In the basket-type centrifuge, flow enters the machine at the bottom and is directed toward the outer wall of the basket. Cake continually builds up within the basket until the centrate, which overflows a weir at the top of the unit, begins to increase in solids. At that point, feed to the unit is shut off, the machine decelerates, and a skimmer enters the bowl to remove the liquid layer remaining in the unit. A knife is then moved into the bowl to cut out the cake, which falls out of the open bottom of the machine. The unit is a batch device with alternate charging of feed sludge and discharging of dewatered cake.

Equipment normally associated with the centrifugal dewatering process includes centrifuge, sludge feed pump, solids conveyor, and centrate pumps. A flow diagram of a centrifugal dewatering system is shown below.



(SEE SECTION IV.7)

IV.6.12.2 Typical Design Criteria

Each installation is site specific and dependent upon a manufacturer's product line. Maximum capacities of about 100 tons/hr of dry solids are available in solid-bowl units with diameters up to 54 inches and power requirements up to 175 hp. Disc units are available with capacities up to 400 gpm of concentrate.

IV.6.12.3 Costs

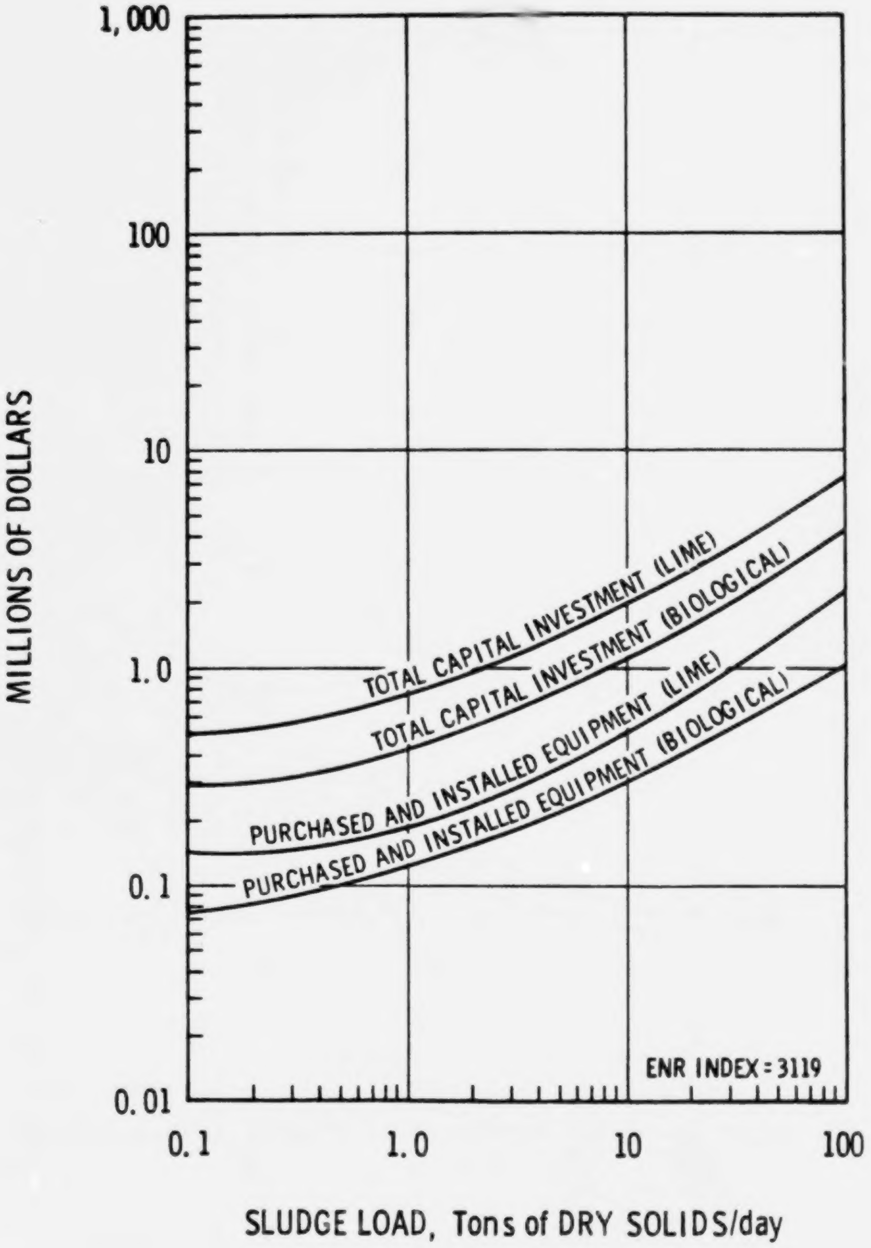
Purchased equipment and installation cost for estimation of the total capital investment includes centrifuges (solid bowl), with minimum of one spare; sludge pumps and piping; cake conveyors; and internal electrical and building costs. Costs do not include centrate handling. Cationic polymer cost is based on 10 lb/ton dry basis for biological sludge. The following operating characteristics were assumed:

Operating characteristic	Assumed value
Sludge quantity	4,500 lb/Mgal at 10% solids for lime sludge; 900 lb/Mgal at 4% for digested biological sludge
Operation	8 hr/d

IV.6.12.4 References

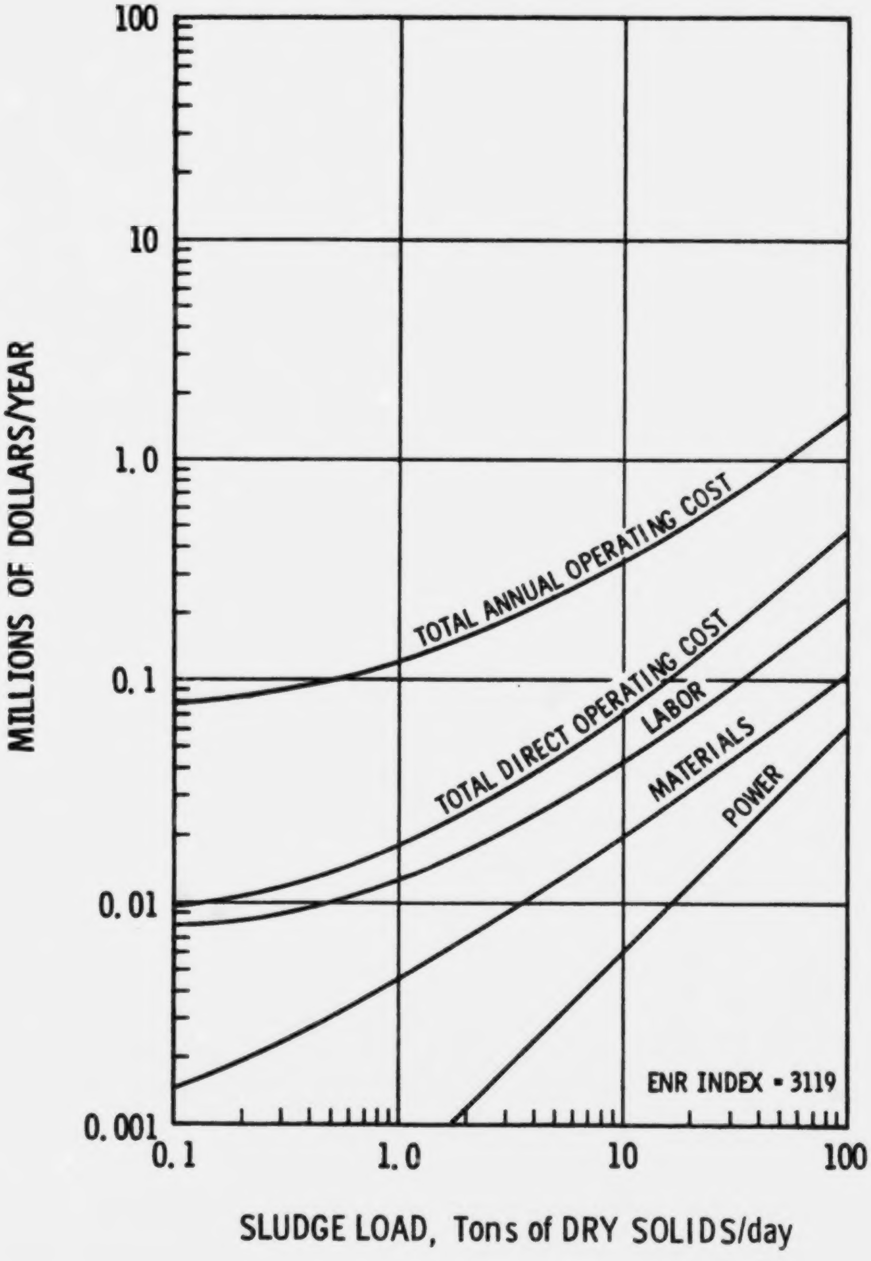
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

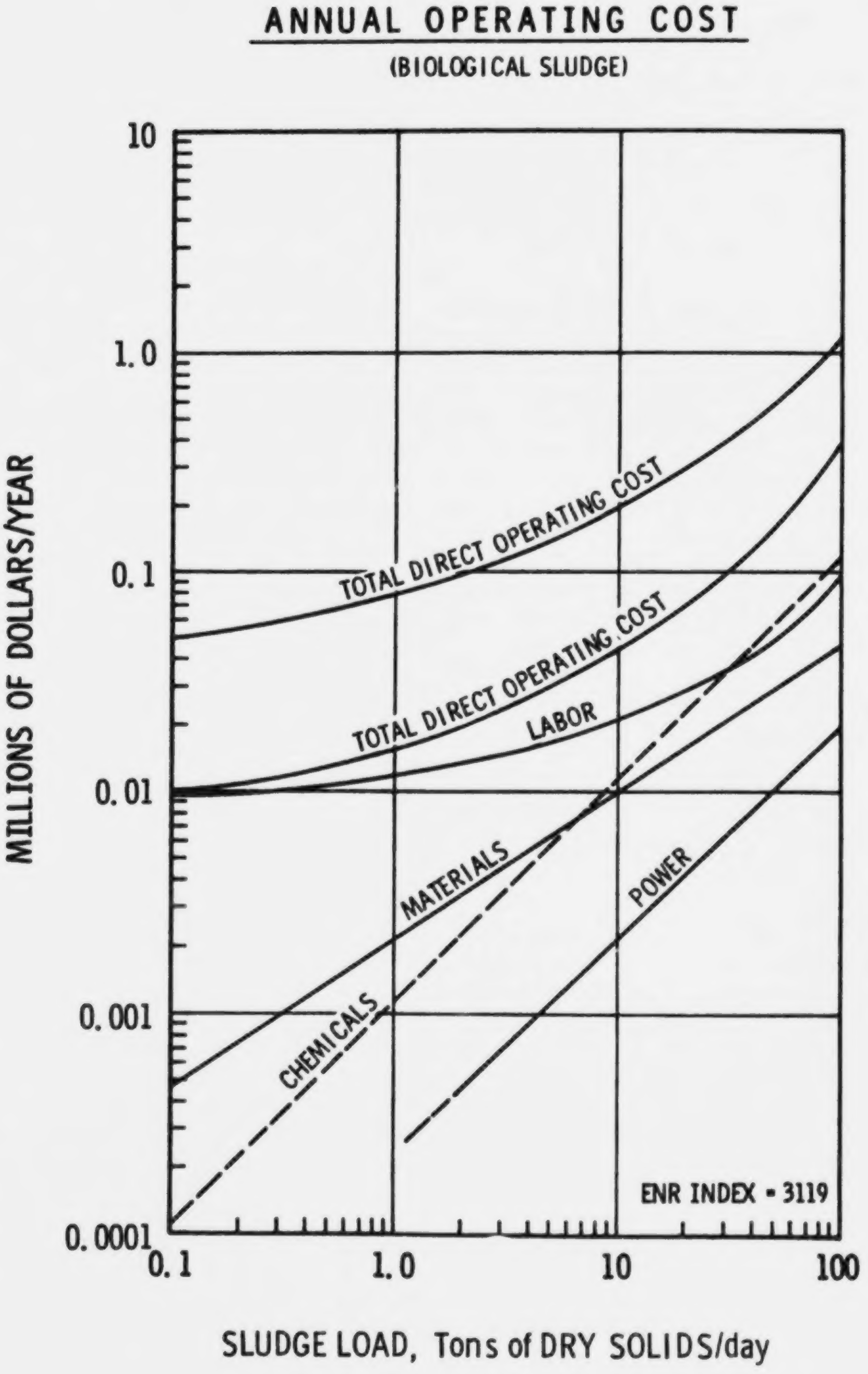
TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST

(LIME SLUDGE)




ANNUAL OPERATING COST
(BIOLOGICAL SLUDGE)
MILLIONS OF DOLLARS/YEAR
10
1.0
0.1
0.01
0.001
0.0001
TOTAL DIRECT OPERATING COST
TOTAL DIRECT OPERATING COST
LABOR
MATERIALS
CHEMICALS
POWER
ENR INDEX = 3119
0.1
1.0
10
100
SLUDGE LOAD, Tons of DRY SOLIDS/day

IV.6.13 THERMAL DRYING [1]

IV.6.13.1 Description

Thermal drying is the process of reducing the moisture in sludge by evaporation to 8 to 10 percent using hot air, without combusting the solid materials. For economic reasons, the moisture content of the sludge must be reduced as much as possible through mechanical means prior to heat drying. The five available heat treating techniques are flash, rotary, toroidal, multiple hearth and atomizing spray.

Flash drying is the instantaneous vaporization of moisture from solids by introducing the sludge into a hot gas stream. The system is based on several distinct cycles that can be adjusted for different drying arrangements. The wet sludge cake is first blended with some previously dried sludge in a mixer to improve pneumatic conveyance. Blended sludge and hot gases from the furnace at about 1200°F to 1400°F (650 to 760°C) are mixed and fed into a cage mill in which the mixture is agitated and the water vapor flashed. Residence time in the cage mill is only a matter of seconds. Dry sludge with eight-to-ten percent moisture is separated from the spent drying gases in a cyclone, with part of it recycled with incoming wet sludge cake and another part screened and sent to storage.

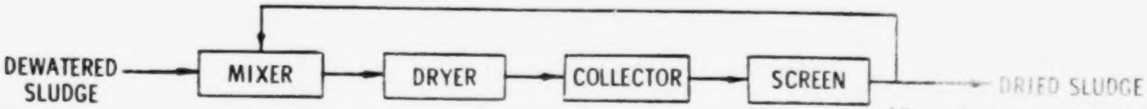
A rotary dryer consists of a cylinder that is slightly inclined from the horizontal and revolves at about five-to-eight r/min. The inside of the dryer is equipped usually with flights or baffles throughout its length to break up the sludge. Wet cake is mixed with previously heat dried sludge in a pug mill. The system may include cyclones for sludge and gas separation, dust collection scrubbers, and a gas incineration step.

The toroidal dryer uses the jet mill principle, which has no moving parts, dries, and classifies sludge solids simultaneously. Dewatered sludge is pumped into a mixer where it is blended with previously dried sludge. Blended material is fed into a doughnut-shaped dryer, where it comes into contact with heated air at a temperature of 800°F to 1100°F. Particles are dried, broken up into fine pieces, and carried out of the dryer by the air stream. The dried, powdered sludge is supplemented with nitrogen and phosphorus and formed into briquettes, which are crushed and screened to produce final products.

The multiple hearth furnace is adapted for heat drying of sludge by incorporating fuel burners at the top and bottom hearths, plus down draft of the gases. The dewatered sludge cake is mixed in a pug mill with previously dried sludges before entering the furnace. At the point of exit from the furnace, the solids temperature is about 100°F, and the gas temperature is about 325°F.

Atomizing drying involves spraying liquid sludge in a vertical tower through which hot gases pass downward. Dust carried with hot gases is removed by a wet scrubber or dry dust collector. A high-speed centrifugal bowl can also be used to atomize the liquid sludge into fine particles and to spray them into the top of the drying chamber where moisture is transferred to the hot gases.

Complete heat drying systems are generally proprietary. The major equipment includes mixers, furnaces, cyclones, screens, dryers, wet scrubbers, dust collectors, air blowers, heaters, spraying devices, sludge feed pumps, and handling equipment. A flow diagram of the thermal drying process is shown below.



(See Section IV.7)

IV.6.13.2 Typical Design Criteria

Approximately 1,400 Btu are needed to vaporize one pound of water, based on a thermal efficiency of 72 percent. Less fuel would be required with additional heat recovery. Chemical scrubbers are used, or chemicals are added prior to heat drying. Excessive drying tends to produce a sludge that is dusty or contains many fine particles; this is less acceptable for marketing and should be avoided. Wet scrubbers and/or solids collectors are needed. Standby heat-drying equipment is needed for continuous operation.

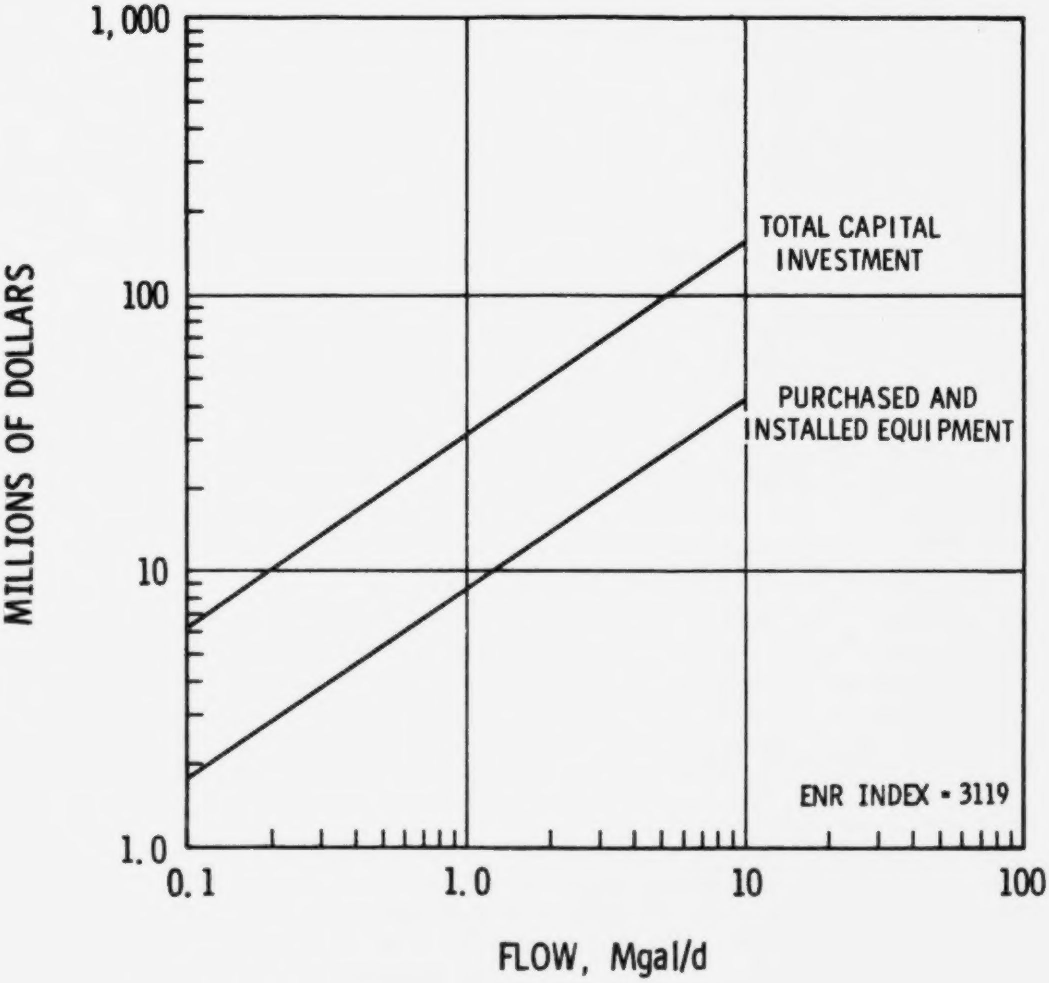
IV.6.13.3 Costs

The costs of heat treatment are based upon the installation of a Zimpro unit and include a complete oxidation unit, installation, building and foundation and piping for the removal of effluent. The equipment is sized to treat a three-percent solids content sludge with a continuous 24-hour-per-day operation, allowing for a 20-percent down time for maintenance.

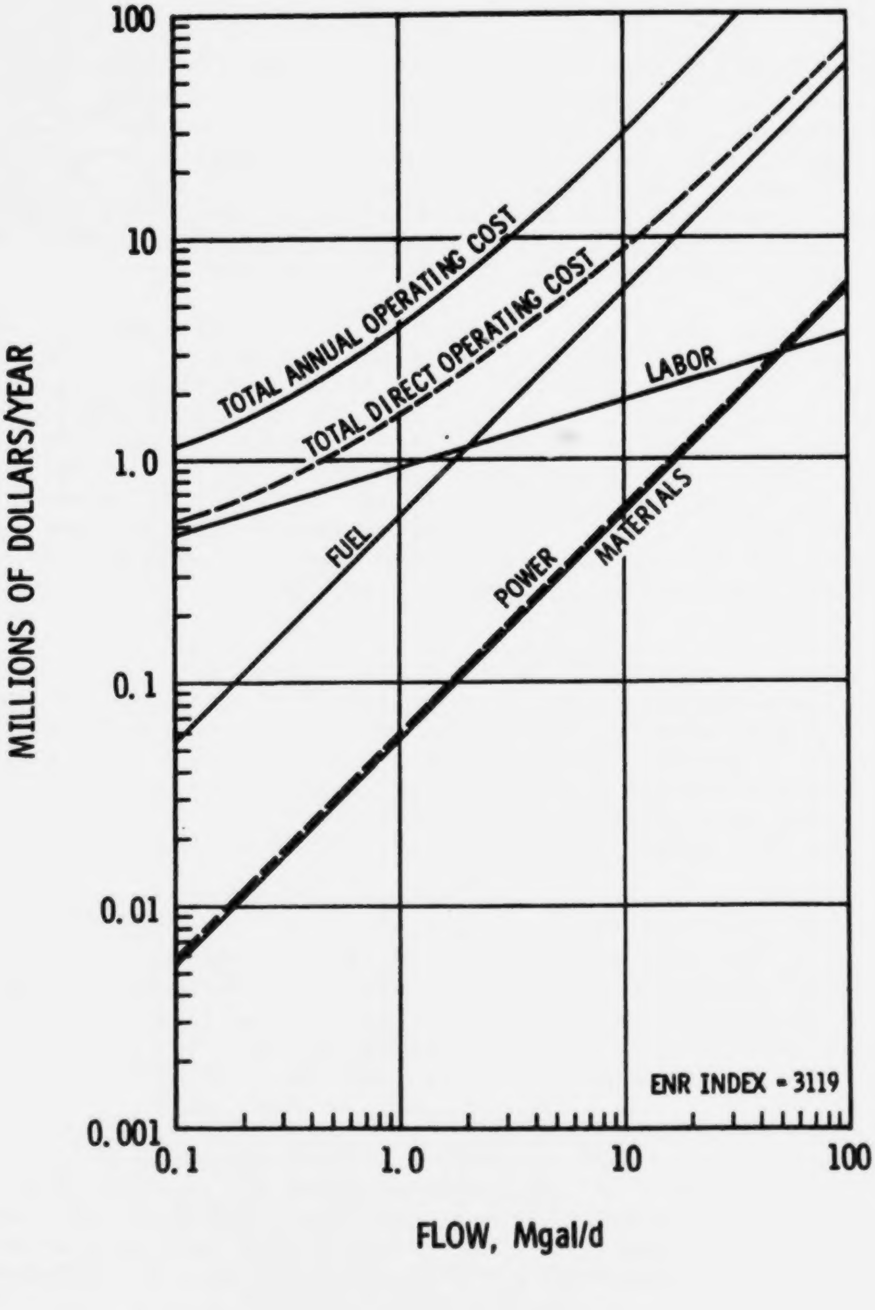
IV.6.13.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.6.14 DRYING BEDS [1]

IV.6.14.1 Description

Drying beds are used to dewater sludge both by drainage through the sludge mass and by evaporation from the surface exposed to the air. Collected filtrate is usually returned to the treatment plant.

Drying beds usually consist of 4 to 9 inches of sand, which is placed over 8 to 18 inches of graded gravel or stone. The sand typically has an effective size of 0.3 to 1.2 mm and a uniformity coefficient of less than 5.0. Gravel is normally graded from 1/8 to 1.0 inch. Drying beds have underdrains that are spaced from 8 to 20 feet apart. Underdrain piping is often vitrified clay laid with open joints and having a minimum diameter of 4 inches and a minimum slope of about 1 percent.

Sludge is placed on the beds in an 8- to 12-inch layer. The drying area is partitioned into individual beds, approximately 20 feet wide by 20 to 100 feet long, of a convenient size so that one or two beds will be filled by a normal withdrawal of sludge from the digesters. The interior partitions commonly consist of two or three creosoted planks, one on top of the other, to a height of 15 to 18 inches, stretching between slots in precast concrete posts. The outer boundaries may be of similar construction or earthen embankments for open beds, but concrete foundation walls are required if the beds are to be covered.

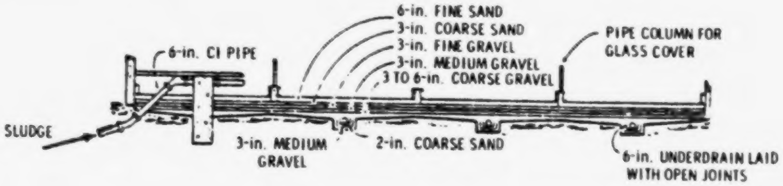
Piping to the sludge beds is generally made of cast iron and designed for a minimum velocity of 2.5 feet/second. It is arranged to drain into the beds and provisions are made to flush the lines and prevent freezing in cold climates. Distribution boxes are provided to divert sludge flow to the selected bed. Splash plates are used at the sludge inlets to distribute the sludge over the bed and prevent erosion of the sand.

Sludge can be removed from the drying bed after it has drained and dried sufficiently to be spadable. Sludge removal is accomplished by manual shoveling into wheelbarrows or trucks, or by a scraper or front-end loader. Provisions should be made for driving a truck onto or along the bed to facilitate loading. Mechanical devices can remove sludges of 20% to 30% solids while cakes of 30% to 40% are generally required for hand removal.

Paved drying beds with limited drainage systems permit the use of mechanical equipment for cleaning. Field experience indicates that the use of paved drying beds results in shorter drying times as well as more economical operation when compared with conventional sandbeds because, as indicated above, the use of mechanical equipment for cleaning permits the removal of sludge with a higher moisture content than does hand cleaning. Paved beds have worked successfully with anaerobically digested sludges but are less desirable than sandbeds for aerobically digested activated sludge.

Sandbeds can be enclosed by glass. Glass enclosures (1) protect the drying sludge from rain, (2) control odors and insects, (3) reduce the drying periods during cold weather, and (4) can improve the appearance of a waste treatment plant.

Equipment normally associated with drying bed operations includes front-end loaders and scrapers. A flow diagram of a drying bed is shown below.


6-in. FINE SAND
3-in. COARSE SAND
3-in. FINE GRAVEL
3-in. MEDIUM GRAVEL
3 TO 6-in. COARSE GRAVEL
PIPE COLUMN FOR GLASS COVER
6-in. CI PIPE
SLUDGE
3-in. MEDIUM GRAVEL
2-in. COARSE SAND
6-in. UNDERDRAIN LAID WITH OPEN JOINTS

IV.6.14.2 Typical Design Criteria

Type of sludge	Area of drying beds sq ft/capita	
	Open beds	Covered beds
Primary digested	1.0 to 1.5	0.75 to 1.0
Primary and humus digested	1.25 to 1.75	1.0 to 1.25
Primary and activated digested	1.75 to 2.5	1.25 to 1.5
Primary and chemically precipitated digested	2.0 to 2.5	1.25 to 1.5

In the southern United States, reduced areas are often practical because of more favorable climatic conditions.

Enclosed beds usually require 60% to 75% of the open bed area. Solids loading rates vary from 10 to 28 $lb/ft^2/yr$ for open beds and 12 to 40 $lb/ft^2/yr$ for closed beds. Sludge beds should be located at least 200 feet from dwellings to avoid odor complaints due to poorly digested sludges.

IV.6.14.3 Costs

Purchased equipment and installation cost for estimation of total capital investment includes sand beds, sludge inlets, underdrains, cell dividers, sludge piping, underdrain return, and other structural elements of the beds. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Service life	20 yr
Bed loading	900 lb sludge/Mgal; 20 lb/ft²/yr

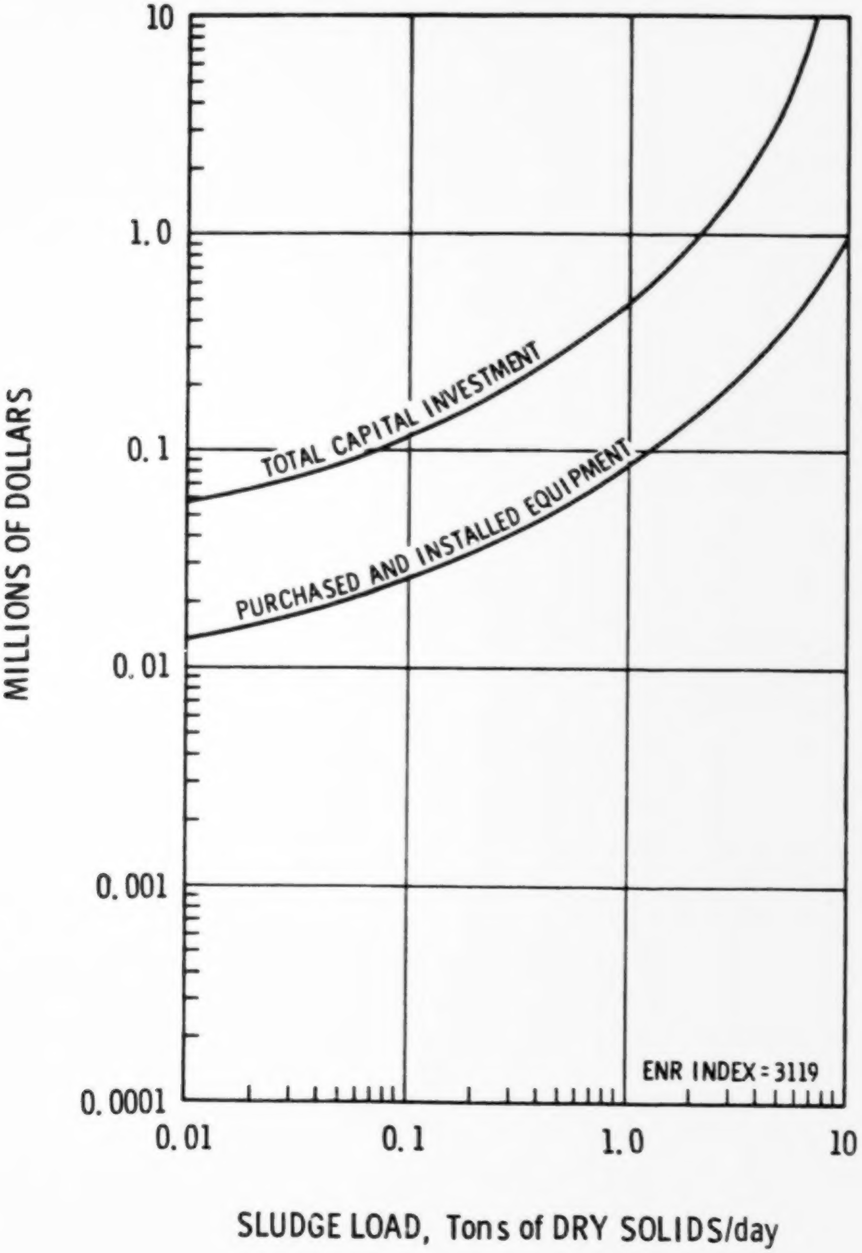
To adjust costs for alternative bed loading rates, sludge quantities, or characteristics, calculate the effective loading (L_E) using the following equation:

$$L_E = L_{Design} \times \frac{20 \text{ lb/ft}^2\text{/yr}}{\text{new bed loading}}$$

IV.6.14.4 References

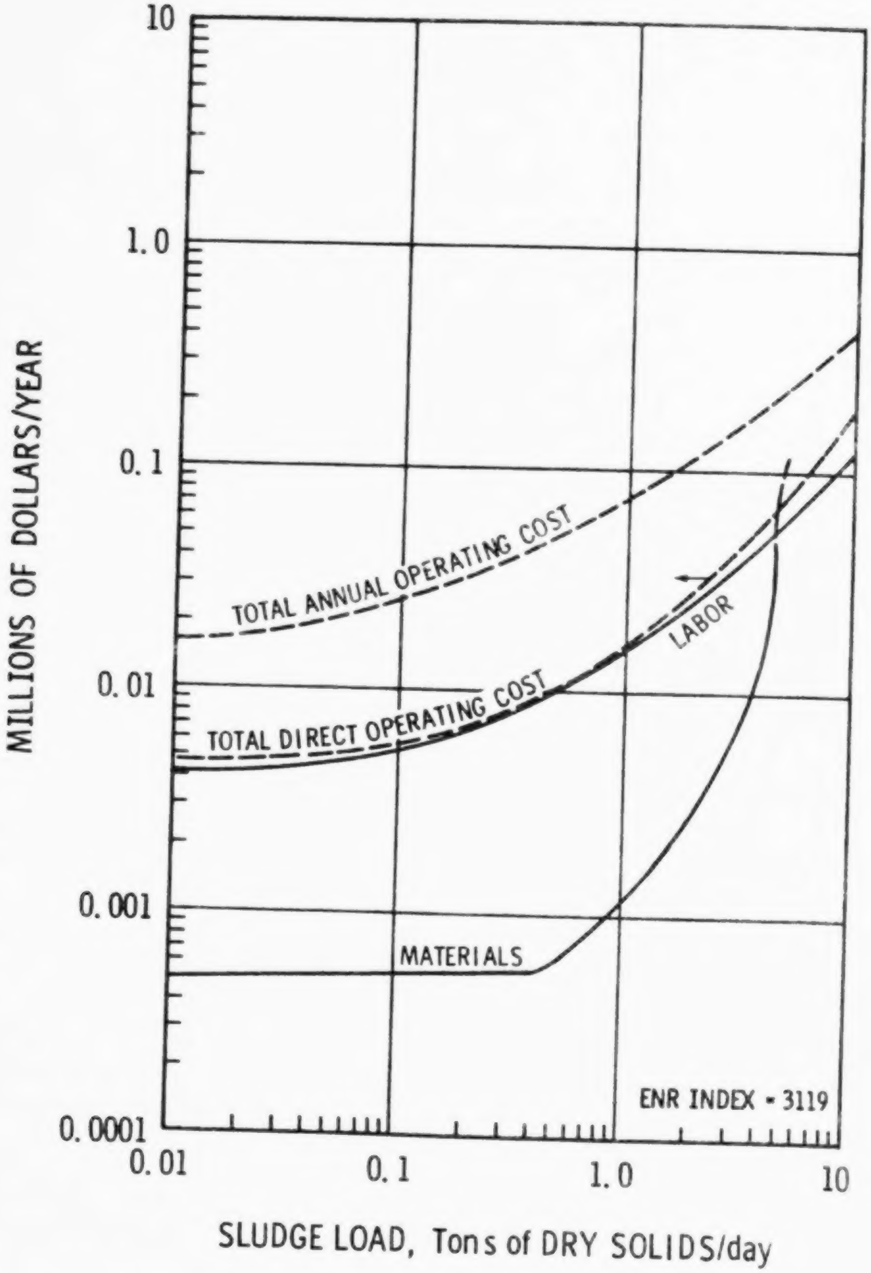
1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



Date: 6/23/80

ANNUAL OPERATING COST



IV.6.15 SLUDGE LAGOONS [1]

IV.6.15.1 Description

Digested sludge has often been applied to sludge lagoons adjacent to or in the proximity of treatment facilities. These sludge lagoons are primarily designed to accomplish long-term drying of the digested sludge through the physical processes of percolation and evaporation, primarily the latter.

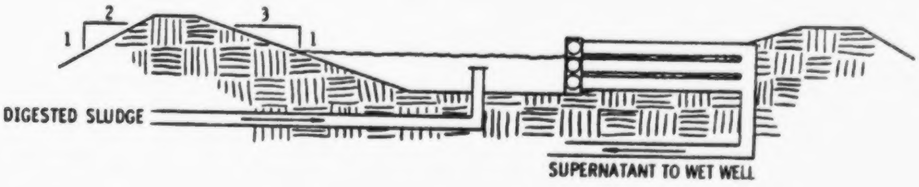
This method of sludge processing has been extremely popular in the U.S. due to its relatively low cost (when inexpensive land is plentiful) and minimal operation and maintenance requirements, especially at smaller wastewater treatment facilities. The process is relatively simple, requiring periodic decanting of supernatant back to the head of the plant and occasional mechanical excavation of dewatered or dried sludge for transportation to its ultimate disposal location. Lagoons can be a very useful process step. Lagoon supernatant is far better (low SS) than supernatant from a secondary digester or even a thickener. Ultimate disposal of the product solids often is as a soil conditioner or landfill.

Sludge lagoons may also be used as contingency units at treatment plants to store and/or process sludges when normal processing units are either overloaded or out of service.

The drying time to 30% solids is generally quite lengthy and may require years. Climatic conditions and pre-lagoon sludge processing greatly influence lagoon performance. In warmer, drier climates well-digested sludges are economically and satisfactorily treated by sludge-drying lagoons because of their inherent simplicity of operation and flexibility. Complete freezing causes sludge to agglomerate; hence, when it thaws, the supernatant decants or drains away easily. Well-digested sludges minimize potential odor problems and are inherent in this type of system. Multiple-cells are required for efficient operation.

Methods and patterns of loading, supernatant recycling techniques, and mechanical cleaning techniques vary with location, climate, and type of sludge to be processed.

Equipment normally associated with sludge lagooning includes front-end loaders, bulldozers, and dragline. A flow diagram of a sludge lagoon is shown below.



IV.6.15.2 Typical Design Criteria

Typical design criteria for sludge lagoons are shown below.

Design criteria	Value
Dikes	Slopes of 1:2 exterior and 1:3 interior are needed to permit maintenance and mowing and to prevent erosion; width must be sufficient to allow vehicle transport during cleaning.
Depth	1.5 to 4.0 ft of sludge depth (depending on climate).
Bottom	Separation from groundwater is dependent upon application depths and soil characteristics, but should not be less than 4 ft to prevent groundwater contamination. An impermeable liner may be required.
Cells	A minimum of two cells is required.
Loading rates	2.2 to 2.4 lb solids/yr/ft^3 of capacity; 1.7 to 3.3 lb solids/ft^2 of surface/30 days of bed use.
Decant	Single or multiple-level decant for periodic returning supernatant to head of plant.
Sludge removal	Approximately 1.5 to 3 yr intervals.

IV.6.15.3 Costs

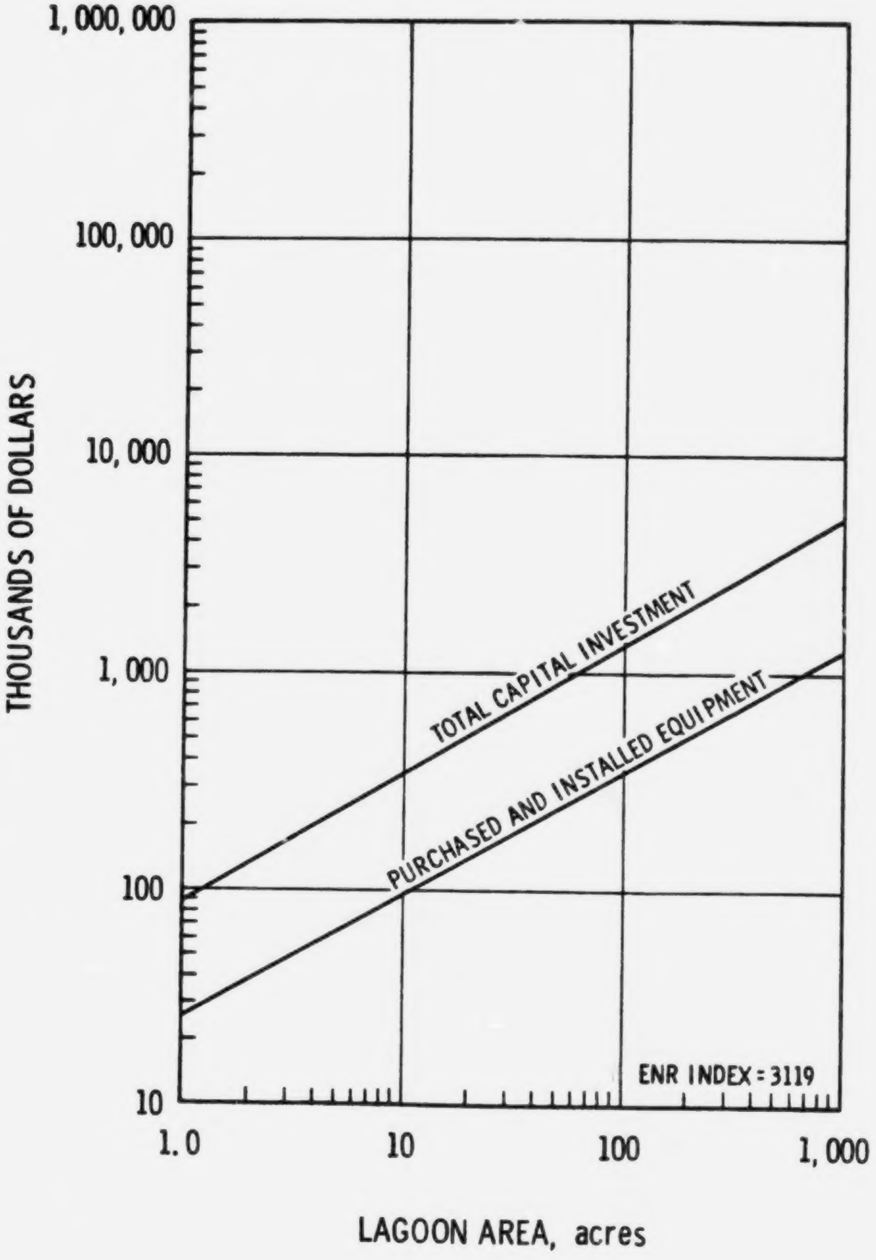
Purchased equipment and installation cost for estimation of total capital investment includes process piping, equipment, concrete, steel, and excavation. The following operating characteristics were assumed for cost estimation:

Operating characteristic	Assumed value
Sizing	75 tons/acre/yr; 1.5 ft depth of sludge
Total annual direct operating costs	Include materials, supplies, maintenance, operation, and residuals removal.

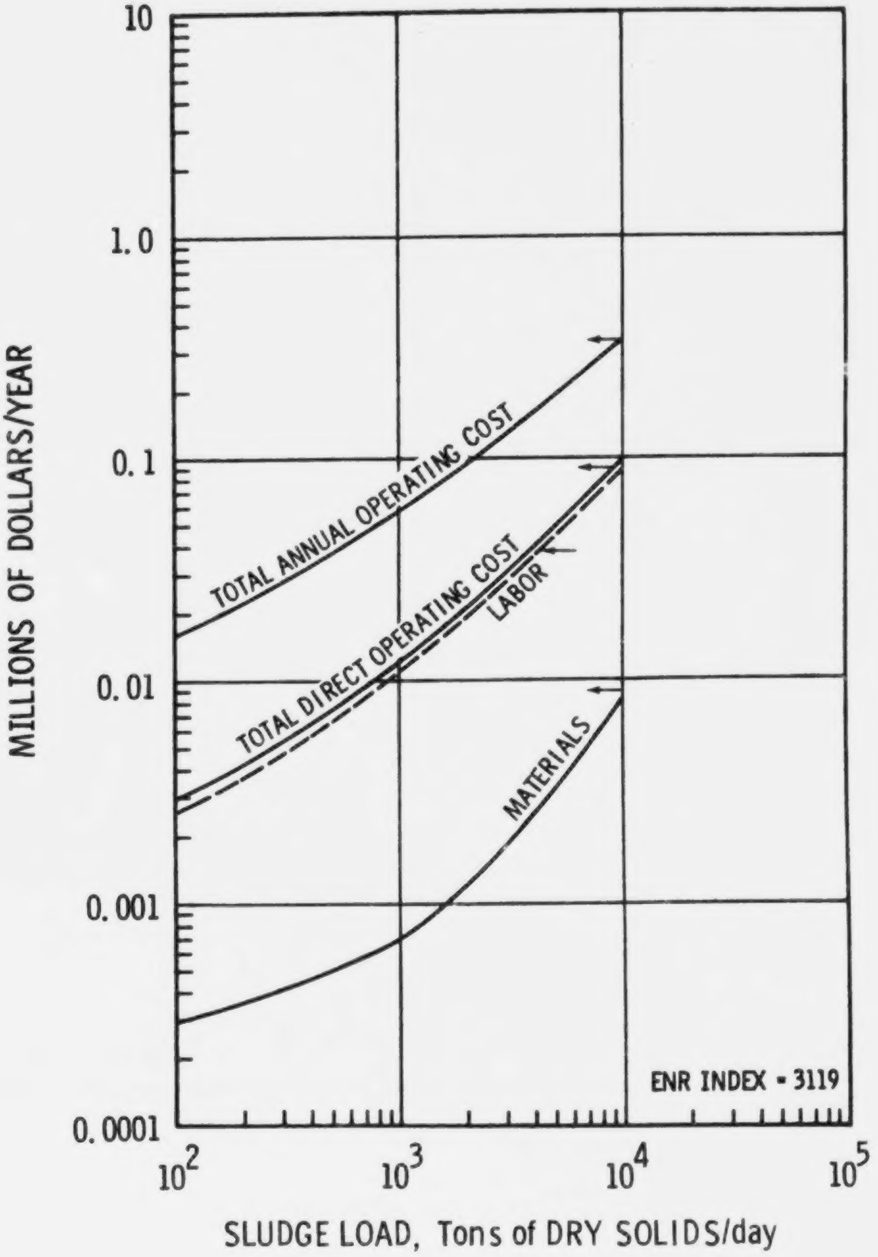
IV.6.15.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7. DISPOSAL

IV.7.1 INCINERATION

IV.7.1.1 Incineration of Sludge - Fluidized Bed Furnace (FBF) [1]

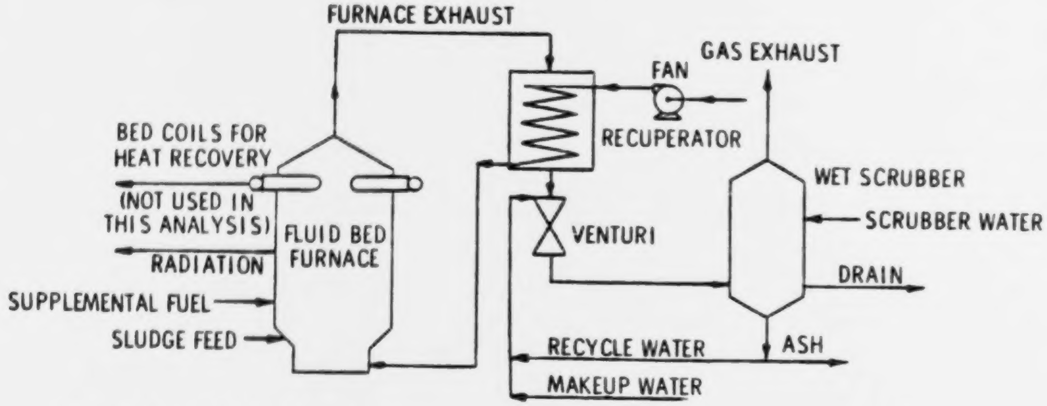
Description. Sludge incineration is a two-step process involving drying and combustion after preliminary dewatering. A typical sludge contains 75% water and 75% volatiles in dry solids. Self-sustained combustion without supplementary fuel is often possible with dewatered raw sludges having a solids concentration greater than 30%.

The fluidized bed furnace (FBF) is a vertically oriented, cylindrically shaped, refractory-lined steel shell that contains a sand bed and fluidizing air distributor. The FBF is normally available in diameters of 9 to 25 feet and heights of 20 to 60 feet. There is one industrial unit operating with a diameter of 53 feet. The sand bed is approximately 2.5 feet thick and rests on a refractory-lined air-distribution grid containing tuyeres through which air is injected at a pressure of 3 to 5 psi to fluidize the bed. Bed expansion is approximately 80% to 100%. Bed temperature is controlled between 1,400°F and 1,500°F by auxiliary burners and/or a water spray or heat removal system above the bed. Ash is carried out the top of the furnace and removed by air pollution control devices, usually wet venturi scrubbers. Sand is lost by attrition at an approximate rate of 5% of the bed volume every 300 hours of operation. Furnace feed can be introduced either above or directly into the bed depending on the type of feed. Generally, sewage sludge is fed directly into the bed.

Excess air requirements for the FBF vary from 20% to 40%. It requires less supplementary fuel than a multiple hearth furnace. An oxygen analyzer in the stack controls the air flow into the reactor, and the auxiliary fuel feed rate is controlled by a bed-temperature controller.

An air preheater is used in conjunction with a fluidized bed to reduce fuel costs. Also, cooling tubes may be submerged in the bed for energy recovery.

Equipment normally associated with sludge incineration includes the FBF, screw pumps, air fans, gas scrubber, and ash handling systems. A flow diagram of a fluidized bed furnace for incineration of sludge is shown on the next page.


FURNACE EXHAUST
GAS EXHAUST
FAN
RECUPERATOR
BED COILS FOR HEAT RECOVERY (NOT USED IN THIS ANALYSIS)
WET SCRUBBER
SCRUBBER WATER
FLUID BED FURNACE
VENTURI
RADIATION
DRAIN
SUPPLEMENTAL FUEL
SLUDGE FEED
RECYCLE WATER
ASH
MAKEUP WATER

Typical Design Criteria. Design criteria for a FBF are shown below. Concerning actual operations, some extensive maintenance problems have occurred with air preheaters. Scaling of the venturi scrubbers has also been a problem. Screw feeds and screw pump feeds are both subject to jamming because of either overdrying of the sludge feed at the incinerator or because of silt carried into the feed system with the sludge. Another frequent problem has been the burnout of spray nozzles or thermocouples in the bed.

Design criteria	Value
Bed loading rate	50 - 60 lb wet solids/ft^2/hr
Superficial bed velocity	0.4 - 0.6 ft/s
Sand effective size	0.2 - 0.3 mm (uniformity coefficient = 1.8)
Operating temperature	1,400 - 1,500¼F (normal); 2,200°F$^+$ (maximum)
Bed expansion	80 - 100%
Sand loss	5% of bed volume per 300 hr of operation

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes the reactor, air blowers, accessories, preheaters, scrubbers, fuel pumps, and building. Costs were based on undigested dewatered primary and secondary sludge (1,900 lb/M gal at 20% solids; 75% volatile).

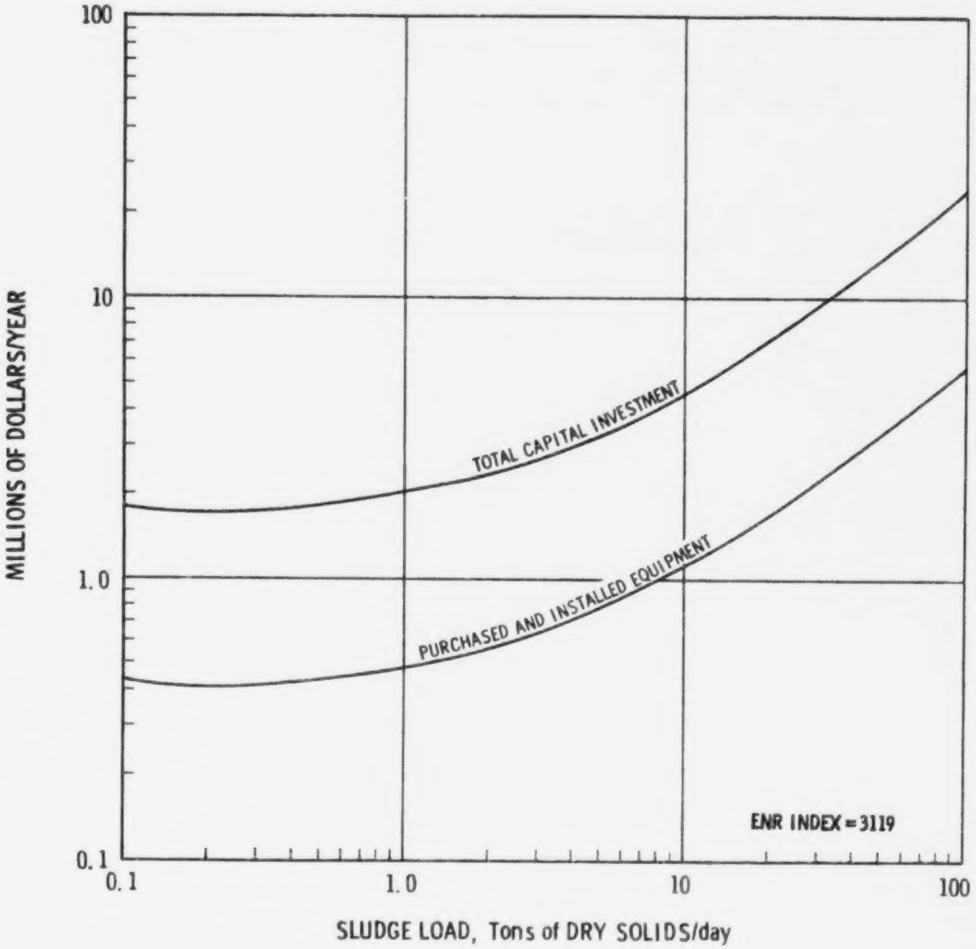
Additionally, the following operating characteristics were assumed.

Plant flow, Mgal/d	Operation, d/wk	Operation, hr/d
0.1	1	20
1.0	7	20
10.0	7	20
100.0	7	20

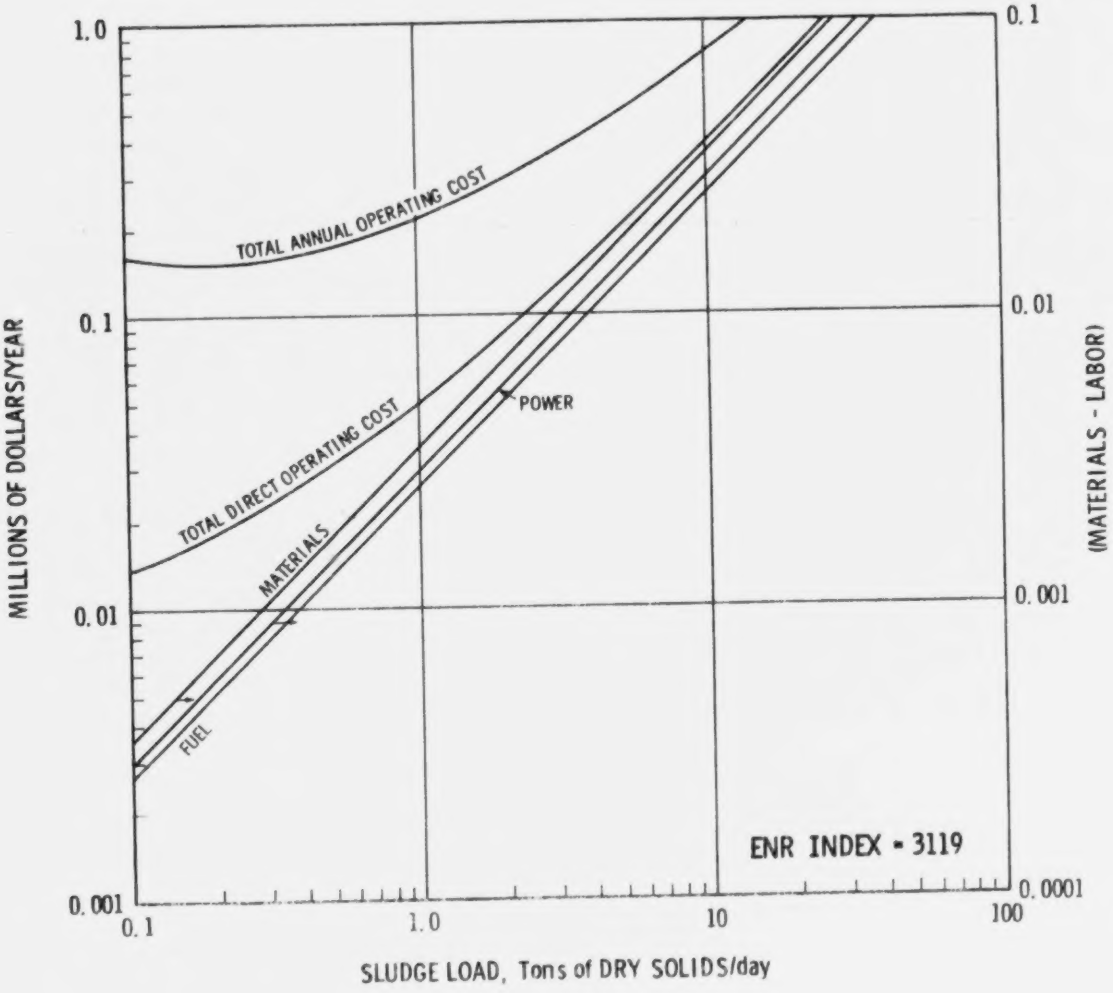
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7.1.2 Incineration of Sludge - Multiple Hearth Furnace (MHF) [1]

Description. Sludge incineration is a two-step process involving drying and combustion after preliminary dewatering. A typical sludge is 80% water and has a dry solids volatility of 75%. Self-sustained combustion without supplementary fuel is often possible with dewatered raw primary sludges, which can frequently be dewatered to 30% solids.

The multiple hearth furnace (MHF) is a vertically oriented, cylindrically shaped, refractory-lined steel shell having diameters of 4 to 25 feet and containing 4 to 13 horizontal hearths positioned one above the other. The hearths are constructed of high heat duty fire brick and special fire brick shapes. Sludge is raked radially across the hearths by rabble arms which are supported by a central rotating shaft that runs the height of the furnace. The cast iron shaft is motor driven with provision for speed adjustment from 1/2 to 1-1/2 r/min. Sludge is fed to the top hearth and proceeds downward through the furnace from hearth to hearth. Inflow hearths have a central port through which sludge passes to the next lower hearth. Outflow hearths have ports on their periphery that also tend to regulate gas velocities. The central shaft contains internal concentric flow passages through which air is routed to cool the shaft and rabble arms. The flow of combustion air is countercurrent to that of the sludge. Gas or oil burners are provided on some hearths for start-up and/or supplemental use as required.

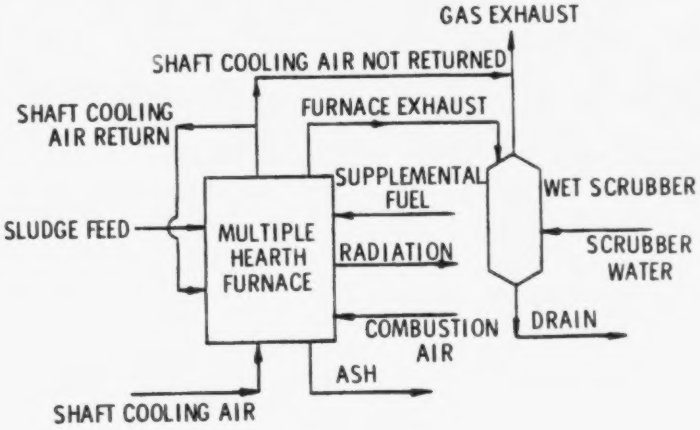
The rabble arms provide mixing action as well as movement to the sludge so that a maximum sludge surface is exposed to the hot furnace gases. Because of the irregular surface left by the rabbling action, the surface area of sludge exposed to the hot gases is as much as 130% of the hearth area. While there is significant solids-gas contact time on the hearths, the overall contact time is actually still greater, due to the fall of the sludge from hearth to hearth through the countercurrent flow of hot gases.

The various phases of the incineration process occur in three zones of the MHF. The drying zone consists of the upper hearths, the combustion zone consists of the central hearths, and the lower hearths comprise the cooling zone. Temperatures in each zone are shown on the following page.

Zone	Temperature, °F Sludge	 Air
Drying	~100	~800
Burning	~1,500	~1,500
Cooling	~400	~350

An afterburner, fired with oil or gas, is provided where required by local air pollution regulations to eliminate unburned hydrocarbons and other combustibles.

Equipment normally associated with multiple hearth furnace incineration of sludge includes the multiple hearth furnace, flapgate valves, cooling and combustion air fans, sludge conveyors, and gas scrubber. A flow diagram of a multiple hearth furnace incineration system is shown below:


GAS EXHAUST
SHAFT COOLING AIR NOT RETURNED
SHAFT COOLING AIR RETURN
FURNACE EXHAUST
SUPPLEMENTAL FUEL
WET SCRUBBER
SLUDGE FEED
MULTIPLE HEARTH FURNACE
RADIATION
SCRUBBER WATER
COMBUSTION AIR
DRAIN
ASH
SHAFT COOLING AIR

Typical design criteria. Typical design criteria for MHF are shown below.

Design criteria	Value
Maximum operating temperature	1,700°F
Hearth loading rate	6 - 10 lb wet solids/ft²/hr with a dry solids concentration of 20 to 40%
Combustion air flow	12 - 13 lb/lb dry solids
Shaft cooling air flow	1/3 - 1/2 of combustion air flow
Excess air	75 - 100%

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes incinerator, building, sludge, conveyor, ash handling equipment, and gas scrubbers. Costs were based on undigested dewatered primary and secondary sludge (1,900 lb/Mgal at 20% solids, 75% volatile). Fuel requirements for warm-up and incineration are 4,500 x 10^6 Btu/yr/Mgal/d.

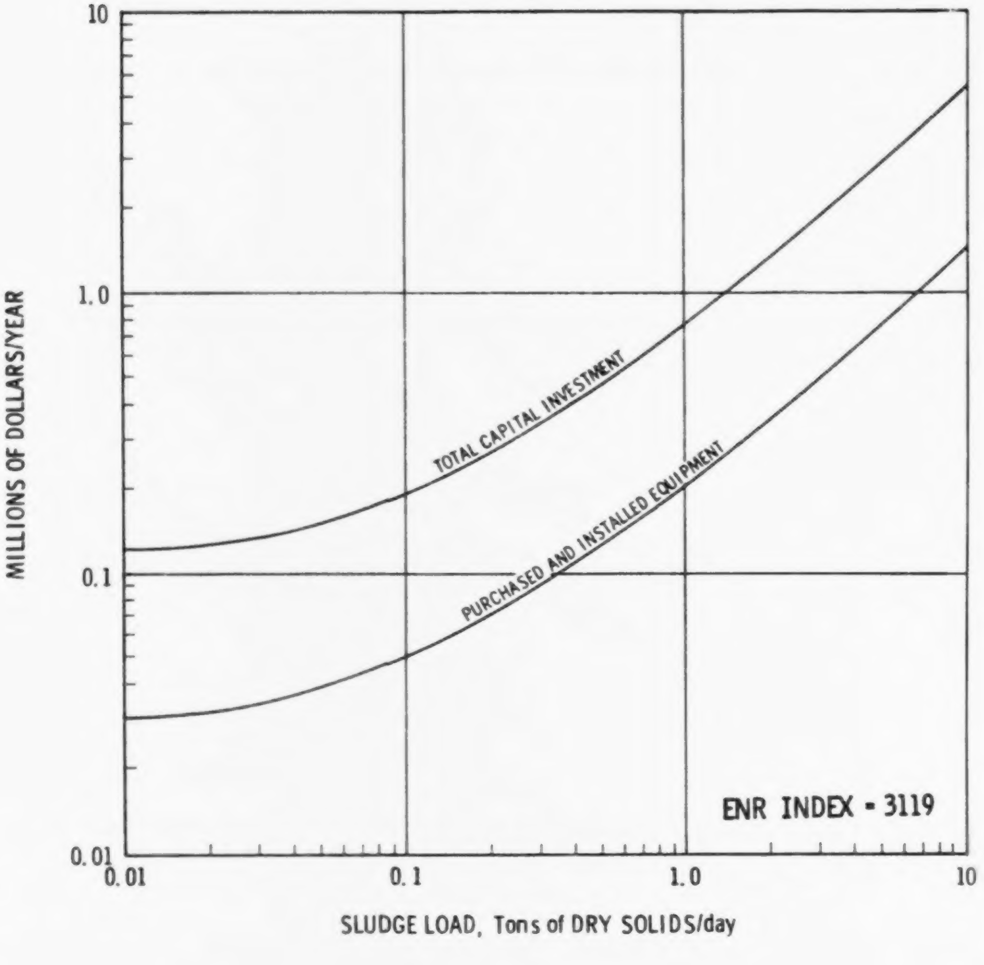
Additionally, the following operating characteristics were assumed:

Plant flow, Mgal/d	Operation, d/wk	Operation, hr/d
0.1	1	20
1.0	7	20
10.0	7	20
100.0	7	20

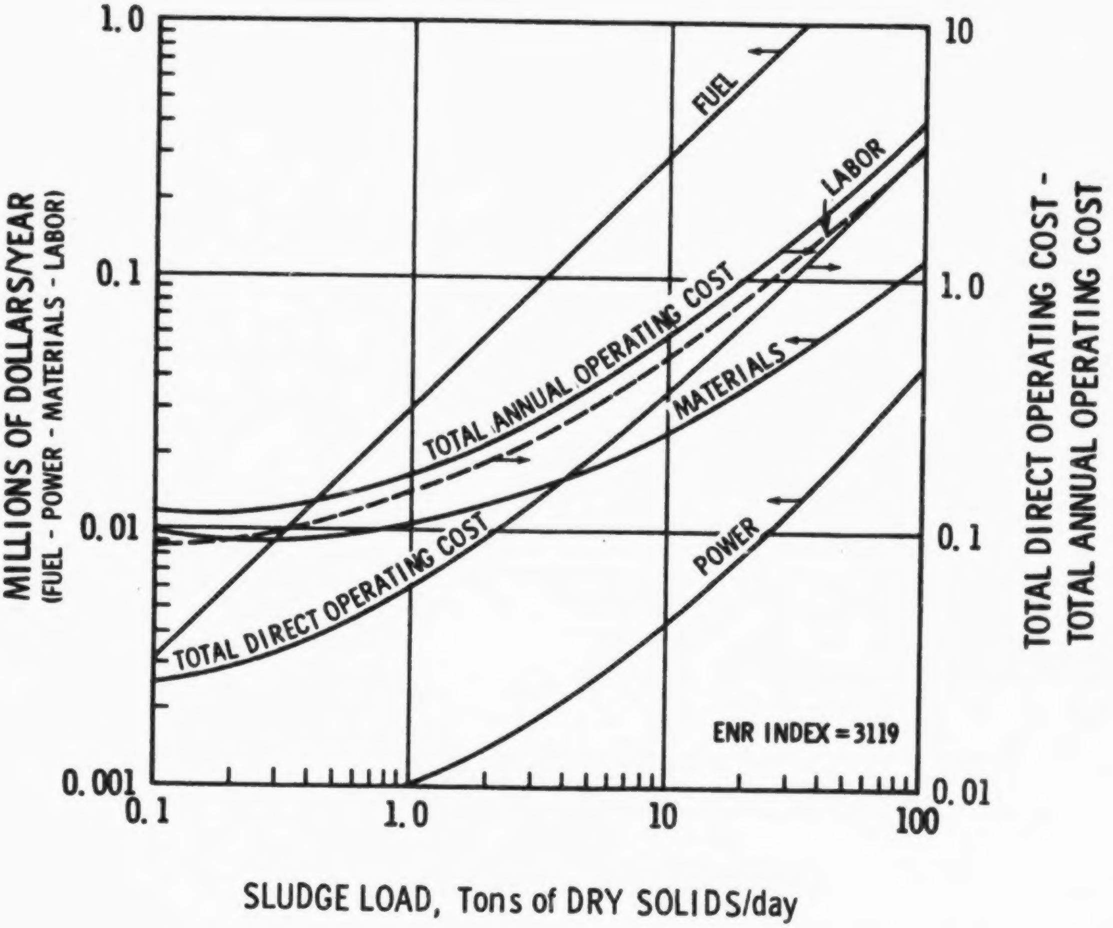
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



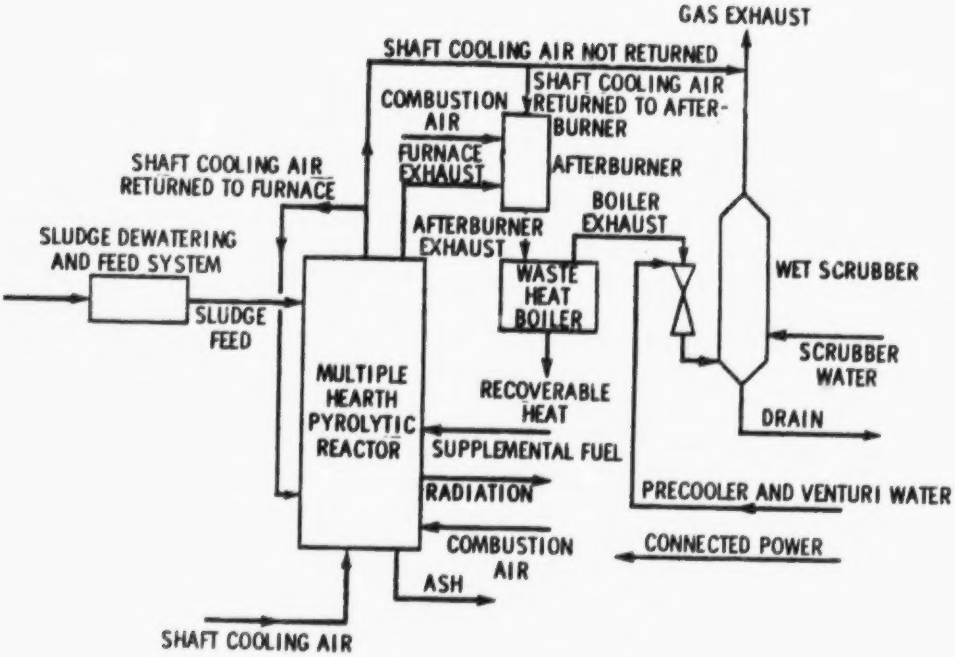
IV.7.2 STARVED AIR COMBUSTION (SAC) [1]

IV.7.2.1 Description

Starved air combustion is used for the volumetric and organic reduction of sludge solids. The process utilizes equipment and process flows similar to incineration except that less than the theoretical amount of air for complete combustion is supplied. Autogenous starved air combustion (SAC) can be achieved with a sludge solids concentration greater than 25%. For lower concentrations, an auxiliary fuel may be required, depending on the percent volatiles in the solids. High temperatures decompose or vaporize the solid components of this sludge. The gas phase reactions are pyrolytic or oxidative, depending on the concentration of oxygen remaining in the stream. Under proper control, the gas leaving the vessel is a low-Btu fuel that can be burned in an after burner to produce power and/or thermal energy. Some processes utilize pure oxygen instead of air and thus produce a higher-Btu fuel gas. The solid residue is a char with more or less residual carbon, depending on how much combustion air had to be supplied to reach the proper operating temperatures. Because the process is neither purely pyrolytic or purely oxidative, it is called starved-air combustion or thermal gasification, rather than pyrolysis. Other processes still in the development stage use indirect heating, rather than the partial combustion. These are true pyrolysis processes. SAC reduces the sludge volumes and sterilizes the end product. Unlike incineration, it offers the potential advantages of producing useful by-products and of reducing the volume of sludge without large amounts of supplementary fuels. The gas that is produced has a heating value of up to 130 Btu/standard dry cubic foot using air for combustion and is suitable for use in local applications, such as combustion in an afterburner or boiler or for fuel in another furnace. SAC has a higher thermal efficiency than incineration due to the lower quantity of air required for the process. In addition, capital economies can be realized due to the smaller gas handling requirements.

Furnaces may be operated in one of three modes resulting in substantially different heat generation and residue characteristics. The low temperature char (LTC) mode only pyrolyzes the volatile material thereby producing a charcoal-like residue with a high ash content. The high temperature char (HTC) mode produces a charcoal-like material converted to fixed carbon and ash. The char burned (CB) mode reacts away all carbon and produces ash as a residue. Heat recovered is maximum for the CB mode, less for the HTC mode, and substantially less for the LTC mode of operation.

Equipment normally associated with the starved air combustion process includes the SAC reactor, waste heat boiler, exhaust gas scrubbers, sludge dewatering devices, and afterburners. A flow diagram is shown on the following page.


GAS EXHAUST
SHAFT COOLING AIR NOT RETURNED
SHAFT COOLING AIR RETURNED TO AFTER-BURNER
COMBUSTION AIR
FURNACE EXHAUST
AFTERBURNER
SHAFT COOLING AIR RETURNED TO FURNACE
BOILER EXHAUST
AFTERBURNER EXHAUST
SLUDGE DEWATERING AND FEED SYSTEM
WASTE HEAT BOILER
WET SCRUBBER
SLUDGE FEED
MULTIPLE HEARTH PYROLYTIC REACTOR
RECOVERABLE HEAT
SCRUBBER WATER
DRAIN
SUPPLEMENTAL FUEL
RADIATION
PRECOOLER AND VENTURI WATER
COMBUSTION AIR
CONNECTED POWER
ASH
SHAFT COOLING AIR

IV.7.2.2 Typical Design Criteria

In MHF systems, hearth loadings are 9 to 15 lb wet (22%) solids/ft²/h; for autogenous combustion, sludge solids content is 25% to 39% depending upon volatility. The off-gas heating value is dependent upon operating mode.

IV.7.2.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes the multiple hearth furnace, fans, motor controls, gas scrubber, external afterburner, ash handling system, auxiliary fuel system, instrumentation, piping, painting, initial operation, and test. The following operating characteristics are assumed for cost estimation:

Operating characteristic	Assumed value
Design capacity	324 ton/d at 40% dry solids
Annual throughput	80,000 tons

Costs are based on third quarter 1979 dollars. Manpower costs were assumed to be $17,500/yr average. Other cost assumptions included: power at $0.02/kWh, fuel at $2.36/10^6 Btu, water and sewer at $0.37/1,000 gal, and residue disposal at $5/ton.

Item	Capital cost
Direct construction cost (DCC)	$2,325,000
Design, construction management (20% DCC)	465,000
Land ($50,000/acre)	250,000
Legal fee (3% DCC)	69,750
Bond discount (3% total cost)	99,000
Total cost	$3,325,000

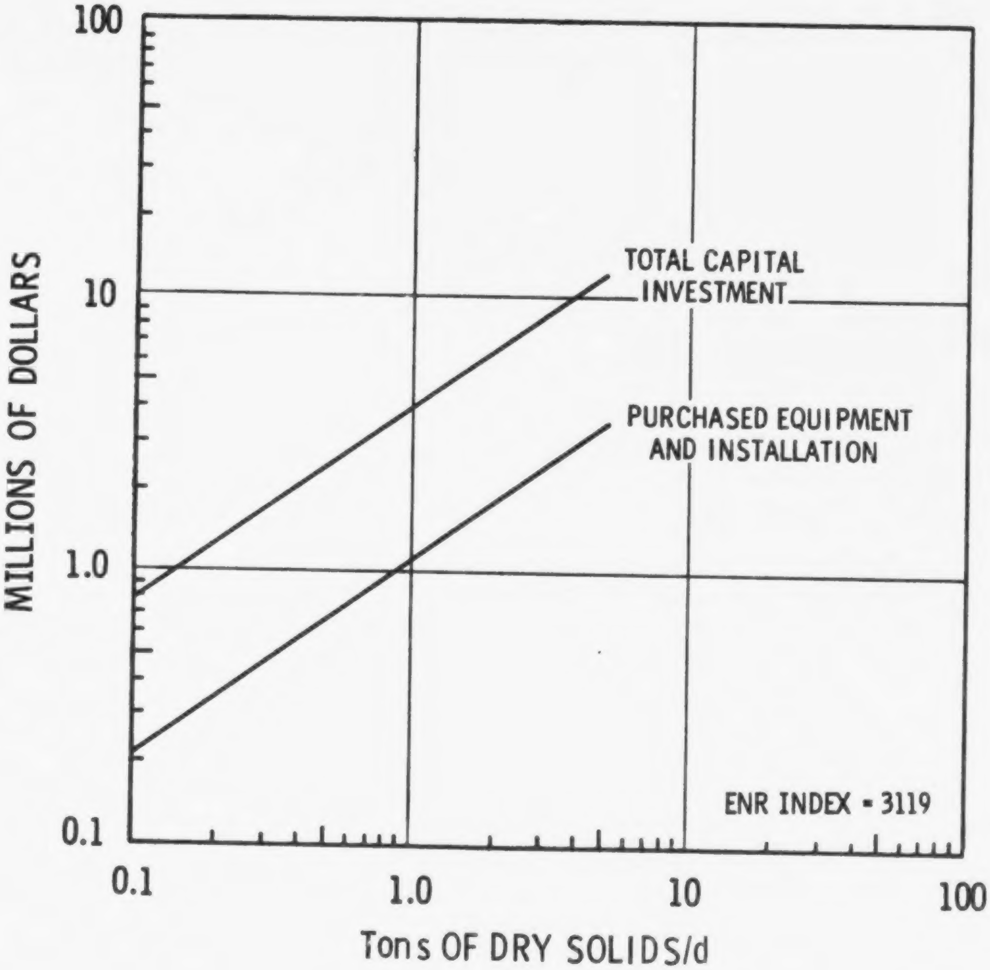
Item	Operating cost Cost/ton[a]	 Annual cost
Manpower, 20 employees	$4.37	$350,000
Power, 210 kWh/hr	0.46	36,800
Water/sewer @ 385 gpm	0.89	70,800
Auxiliary fuel (250,000 gpy)	1.19	95,500
Maintenance (2.5% DCC)	1.03	83,100
Overhead (1% DCC)	0.42	33,250
Residual disposal	0.94	75,000
Total cost	$9.30	$744,450

[a] Based on 80,000 ton/yr throughput.

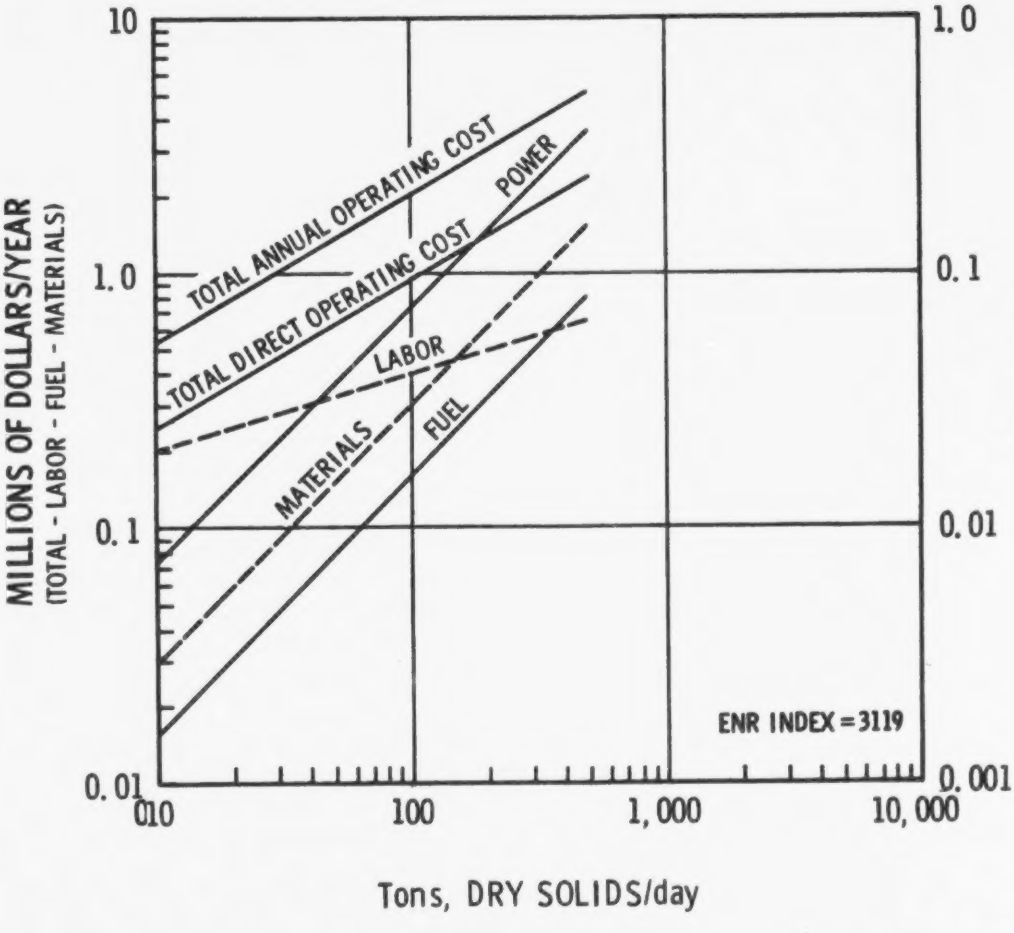
IV.7.2.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7.3 LANDFILLING - AREA FILL [1]

IV.7.3.1 Description

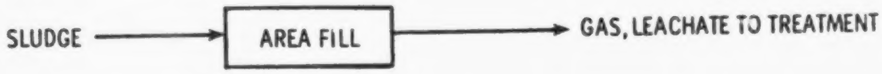
Landfilling is a sludge disposal operation in which sludge is placed above the original earth cover and subsequently covered with soil. To achieve stability and soil-bearing capacity, sludge is mixed with a bulking agent, usually soil. The soil absorbs excess moisture from the sludge and increases its workability. The large quantities of soil required may require hauling from elsewhere. Provisions must be made to keep the stockpiled soil dry. Installation of a liner is generally required for groundwater control and provisions must be made for surface drainage control, gas migration, dust, vectors and/or aesthetics. Area fills are more specifically categorized as follows:

Area Fill Mound. In an area fill mound, sludge is mixed with a bulking agent, usually soil, and the mixture is hauled to the filling area where it is stacked in mounds approximately 6 feet high. Cover material is then applied in a 3-foot thickness. This cover thickness may be increased to 5 feet if additional mounds are applied atop the first lift. The appropriate sludge/soil bulking ratio and soil cover thickness depend upon the solids content of the sludge as received, the need for mound stability, and the bearing capacity as dictated by the number of lifts and equipment weight. Lightweight equipment with swamp pad tracks is appropriate for area fill mound operations; heavier wheel equipment is appropriate in transporting bulking material to and from stockpiles. Construction of earthern containments is useful to minimize mound slumping and for sloping sites.

Area Fill Layer. In an area fill layer, sludge is mixed with soil on or off site and spread evenly in consecutive layers 0.5- to 3-feet thick. Interim cover between layers may be applied to 0.5- to 1-foot thick applications. Layering may continue to an indefinite height before final cover is applied. Lightweight equipment with swamp pad tracks is appropriate for area fill layer operations; heavier wheel equipment is appropriate for hauling soil. Slopes should be relatively flat to prevent sludge from flowing downhill. However, if sludge solids content is high and/or sufficient bulking soil is used, the effect can be prevented and layering performed on mildly sloping terrain.

Diked Containment. Dikes are constructed on level ground around all four sides of a containment area. Alternatively, the containment area may be placed at the toe of the hill so that the steep slope can be utilized as containment on one or two sides. Dikes are then constructed around the remaining sides. Access is provided to the top of the dikes so that haul vehicles can dump sludge directly into the containment. A 1- to 3-foot interim cover may be applied at certain points during the filling; a 3- to 5-foot thick final cover should be applied when filling is discontinued. Cover material is applied either by a dragline, based on solid ground atop the dikes, or by track dozers directly on top of the sludge, depending upon sludge bearing capacity. Usually, operations are conducted without the addition of soil bulking agents, but occasionally soil bulking is added. Typical dimensions are 50 to 100 feet wide, 100 to 200 feet long, 10 to 30 feet deep.

Equipment normally associated with landfilling processes includes front-end loader, bulldozer, scraper, backhoe, dragline, and grader. A flow diagram of landfilling system is shown below:



IV.7.3.2 Typical Design Criteria

Typical design criteria for a sludge landfilling system are listed as follows:

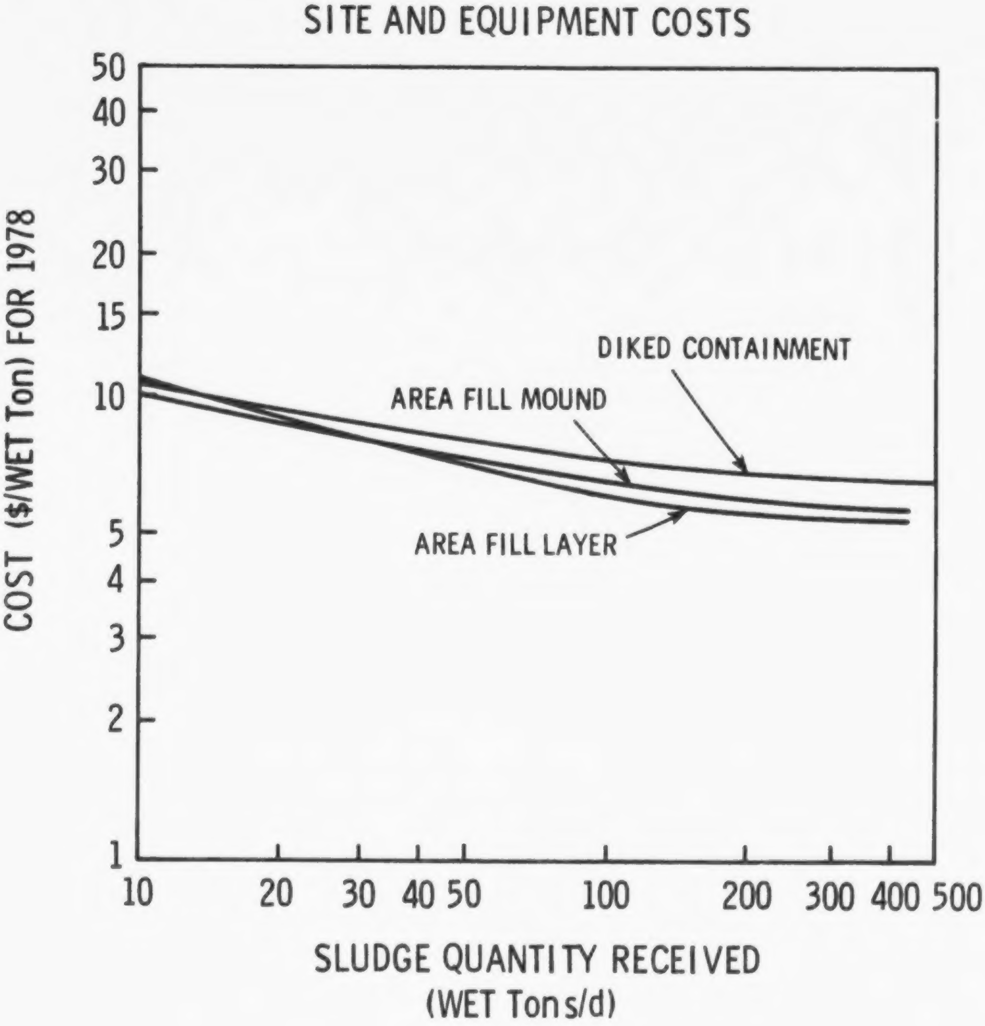
Design criteria	Value		
Parameter	In area fill mound	In area fill layer	In diked containment
Sludge solids content	Greater than 20%	Greater than 15%	20 to 28% for land-based equipment; more than 28% for sludge-based equipment
Sludge characteristics	Stabilized	Stabilized	Stabilized or unstabilized
Ground slopes	No limitation if suitably prepared	Level ground preferred	Level ground or steep terrain if suitably prepared
Bulking required	Yes	Yes	Occasionally
Bulking ratio of soil to sludge	0.5 to 2 parts soil to 1 part sludge	0.25 to 1 part soil to 1 part sludge	0 to 0.5 part soil to 1 part sludge
Sludge application rate	3,000 to 14,000 yd^3/acre	2,000 to 9,000 yd^3/acre	4,800 to 15,000 yd^3/acre
Equipment	Track loader; backhoe with loader; track dozer	Track dozer; grader; track loader	Dragline; track dozer; scraper

IV.7.3.3 Costs

Costs are presented in 1978 dollars. Site and equipment costs include land at $2,500/acre, site preparation (clearing, grubbing, surface water control ditches and ponds, monitoring wells, soil stockpiles, roads and facilities), equipment purchase, and engineering. The actual fill area consumes 50% of the total site area. Operating costs include labor (at $8/hr, including fringe, overhead, and administration), equipment fuel, maintenance and parts, utilities, laboratory analysis of water samples, supplies, and materials. Actual costs vary considerably with specific sludge and site conditions.

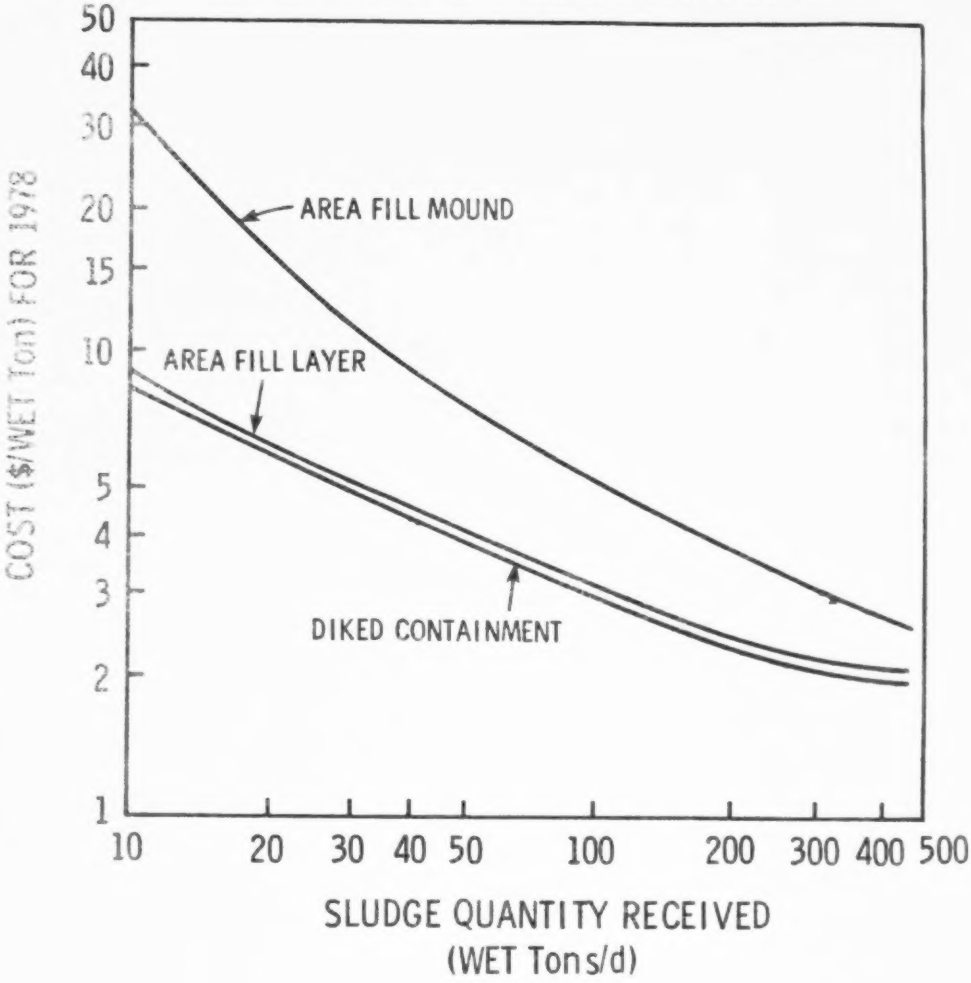
IV.7.3.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-73-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.



SITE AND EQUIPMENT COSTS
COST ($/WET Ton) FOR 1978
50
40
30
20
15
10
5
4
3
2
1
DIKED CONTAINMENT
AREA FILL MOUND
AREA FILL LAYER
10
20
30
40
50
100
200
300
400
500
SLUDGE QUANTITY RECEIVED
(WET Tons/d)

OPERATION AND MAINTENANCE COSTS



IV.7.4 LANDFILLING - SLUDGE TRENCHING [1]

IV.7.4.1 Description

Landfilling is the final disposal of sludge in evacuated trenches. In landfilling, stabilized or unstabilized sludge is placed within a subsurface excavation and covered with soil. Trench operations are more specifically categorized as narrow trench and wide trench. Narrow trenches are defined as having widths less than 10 feet, while wide trenches are defined as having widths greater than that value. The width of the trench is determined by the solids content of the receiving sludge and its capability of supporting cover material and equipment. Distances between trenches should be large enough to provide sidewall stability, as well as space for soil stockpiles, operating equipment, and haul vehicles. Design considerations should include provisions to control leachate and gas migration, dust, vectors, and/or aesthetics. Leachate control measures include the maintenance of 2 to 5 feet of soil thickness between the trench bottom and the highest groundwater level or bedrock (2 feet for clay to 5 feet for sand), or membrane liners and leachate collection and treatment system. Installation of gas control facilities may be necessary if inhabited structures are nearby.

In narrow trench operations, sludge is disposed in a single application, and a single layer of cover soil is applied atop this sludge. Trenches are usually excavated by equipment based on solid ground adjacent to the trench; equipment does not enter the excavation. Backhoes, excavators, and trenching machines are particularly useful. Excavated material is usually applied immediately as cover over an adjacent sludge-filled trench. Sludge is placed in trenches either directly from haul vehicles, through a chute extension, or by pumping. The main advantage of a 2- to 3-foot narrow trench is its ability to handle sludge with a relatively low solids content (15% to 20%). Instead of sinking to the bottom of the sludge, the cover soil bridges over the trench, supported from undisturbed soils along each side of the trench. A 3- to 10-foot width is more appropriate for sludge with solids content of 20% to 28%, which is high enough to support cover soil.

Wide trench operations are usually excavated by equipment operating inside the trench. Track loaders, draglines, scrapers, and track dozers are suitable. Excavated material is stockpiled on solid ground adjacent to the trench for subsequent application as cover materials. If sludge is incapable of supporting equipment, cover is applied by equipment based on solid undisturbed ground adjacent to the trench. A front-end loader is suitable for trenches up to 10 feet wide; a dragline is suitable for trench widths up to 50 feet. If sludge can support equipment, a track dozer applies cover from within the trench. Sludge is placed in trenches from haul vehicles directly entering the trench or from haul vehicles dumping from the top of the trench. Dikes can be used to confine sludge to a specific area in a continuous trench.

After maximum settlement has occurred, in approximately one year, the area should be regraded to ensure proper drainage.

Equipment normally associated with the sludge trenching process includes the front-end loader, bulldozer, scraper, backhoe, dragline, trencher, and grader. A flow diagram of a sludge trenching system is shown below.

SLUDGE
TRENCHING
GAS AND LEACHATE MAY BE COLLECTED AND TREATED

IV.7.4.2 Typical Design Criteria

Typical design criteria for sludge trenching systems are listed below.

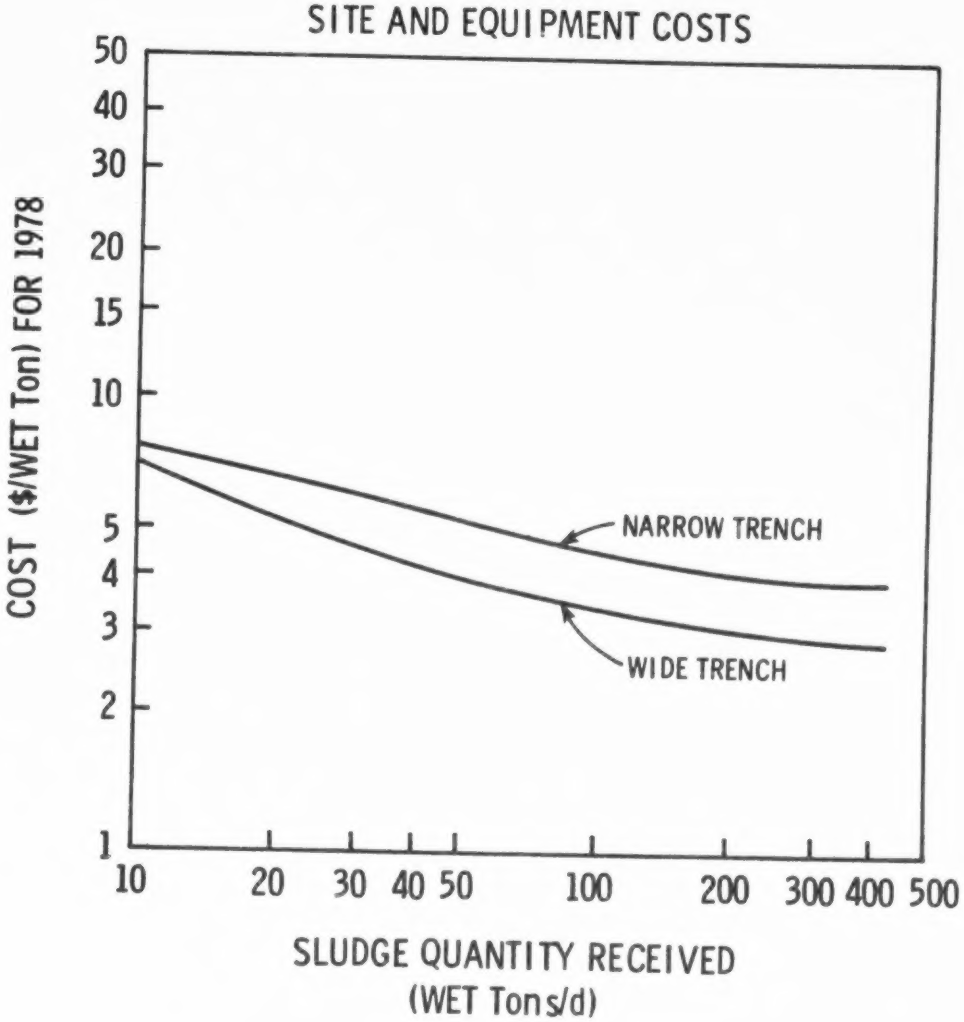
	Value	
Design criteria	Narrow trench	Wide trench
Sludge solids content	15% to 20% for 2- to 3-ft widths; 20% to 28% for 3- to 10-ft widths	20% to 28% for land-based equipment; more than 28% for sludge-based equipment
Ground slopes	Less than 20%	Less than 10%
Cover soil thickness	2 to 3 ft for 2- to 3-ft widths; 3 to 4 ft for 3- to 10-ft widths	3 to 4 ft for land-based equipment; 4 to 5 ft for sludge-based equipment
Sludge application rate	1,200 to 5,600 yd^3/acre	3,200 to 14,500 yd^3/acre
Equipment	Backhoe with loader, excavator, trenching machine	Track loader, dragline, scraper, track dozer

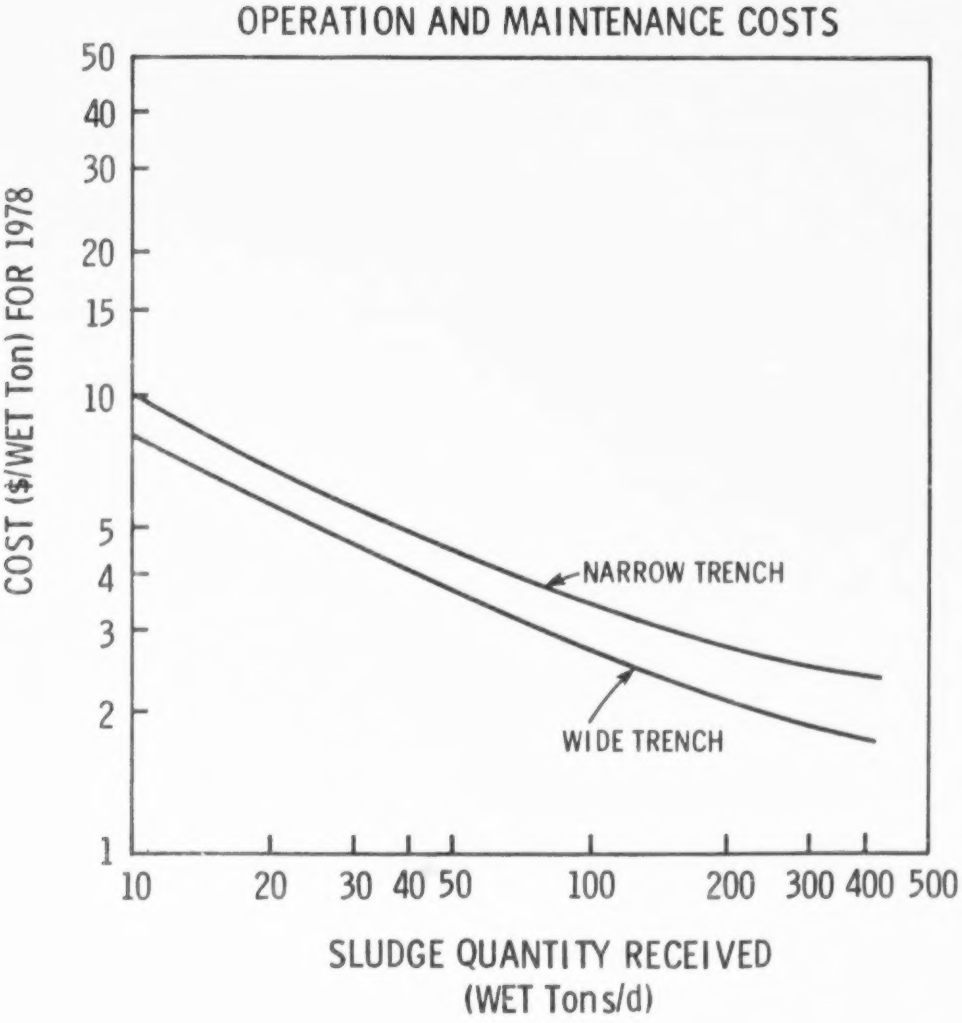
IV.7.4.3 Costs

Site and equipment cost curves, and operation and maintenance cost curves are based on 1978 costs. Site and equipment costs include land at $2,500/acre, site preparation (clearing, grubbing, surface water control ditches and ponds, monitoring wells, soil stockpiles, roads, and facilities), equipment purchase, and engineering. The actual fill area consumes 50% of the total site area. Operating costs include labor (at $8/hr, including fringe, overhead, and administration), equipment fuel, maintenance and parts, utilities, laboratory analysis of water samples, supplies, and materials. Actual cost vary considerably with specific sludge and site conditions.

IV.7.4.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.


SITE AND EQUIPMENT COSTS
COST ($/WET Ton) FOR 1978
50
40
30
20
15
10
5
4
3
2
1
10
20
30
40
50
100
200
300
400
500
NARROW TRENCH
WIDE TRENCH
SLUDGE QUANTITY RECEIVED
(WET Tons/d)


OPERATION AND MAINTENANCE COSTS
COST ($/WET Ton) FOR 1978
50
40
30
20
15
10
5
4
3
2
1
10
20
30
40
50
100
200
300
400
500
SLUDGE QUANTITY RECEIVED
(WET Tons/d)
NARROW TRENCH
WIDE TRENCH

IV.7.5 LAND APPLICATION [1]

IV.7.5.1 Description

Land application techniques for applying liquid sludge, dried sludge, and sludge cake to the land include tank truck, injection, ridge and furrow spreading, and spray irrigation. Sludge can be incorporated into the soil by plowing, discing, or similar methods. Ridge and furrow methods involve spreading sludge in the furrows and planting crops on the ridges. Utilization of this technique is generally best suited to relatively flat land and is well suited to certain row crops. Spray irrigation systems are more flexible, require less soil preparation, and can be used with a wider variety of crops. High application rates are commonly used to reclaim strip mine spoils or other low quality land. Sludge spreading in forests has been limited; however, it offers opportunities for improved soil fertility and increased tree growth.

Farm equipment or tank trucks with standard tires can be adapted for sludge application. However, specially-designed sludge application equipment is now available with high flotation tires and apparatus for applying liquid or dry sludge, or for subsurface injection. This equipment has a 15 psi compaction factor with an 8-ton payload and does minimize rutting, compaction or crop damage when sludge applications are made under proper soil moisture conditions.

IV.7.5.2 Typical Design Criteria

Application rates depend on sludge composition, soil characteristics (usually 3% nitrogen; 2% phosphorus; 0.25% potassium), climate, vegetation, and cropping practices. Annual application rates have varied from 0.5 to more than 100 tons per acre. Applying sludge at a rate to support the nitrogen needs of a crop of about 5 to 10 tons of digested solids in the liquid form, avoids problems associated with overloading the soil. Rates based on phosphorus needs are lower. A pH of 6.5 or greater will minimize heavy metal uptake by most crops.

IV.7.5.3 Costs

Purchased equipment and installation cost for estimation of the total capital investment includes storage lagoon, land preparation, monitoring wells, and service road. Costs for transport of sludge to site and land are not included. The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Service life	30 yr
Storage lagoon	6 wk
Monitoring wells	3 at 0.1 Mgpd 5 at 1 Mgpd 8 at 10 Mgpd 25 at 100 Mgpd
Digested biological sludge	900 lb/Mgal at 4% solids
Sludge application rate	10 ton (dry)/acre/yr
Sludge application	By subsurface injection; unit attached to haul truck

To adjust costs for different application rates, calculate the effective (L_E) using the following equation:

$$L_E = L_{Design} \times \frac{10 \text{ dry ton/acre/yr}}{\text{new design application rate}}$$

An additional example of costs for one project that uses high flotation equipment is provided as follows [1]:

Equipment characteristic	Assumed value
Operation rate	50 wk/yr, 40 hr/wk or 2,000 hr/yr
Application rate	8,000 gal/hr, 5% dry solids = 400 gal dry solids/hr x 8.34 lb/gal = 3,336 lb/hr = 3,336 dry tons/yr
Compaction factor	15 lb/in.^{2} for 8-ton payload
Haul distance, sludge source to spreading site	0.25 mi
Fuel consumption, diesel	6 gal/hr
gas	9 gal/hr
Service life	10 yr

The estimated costs (1977 dollars) are shown on the following page.

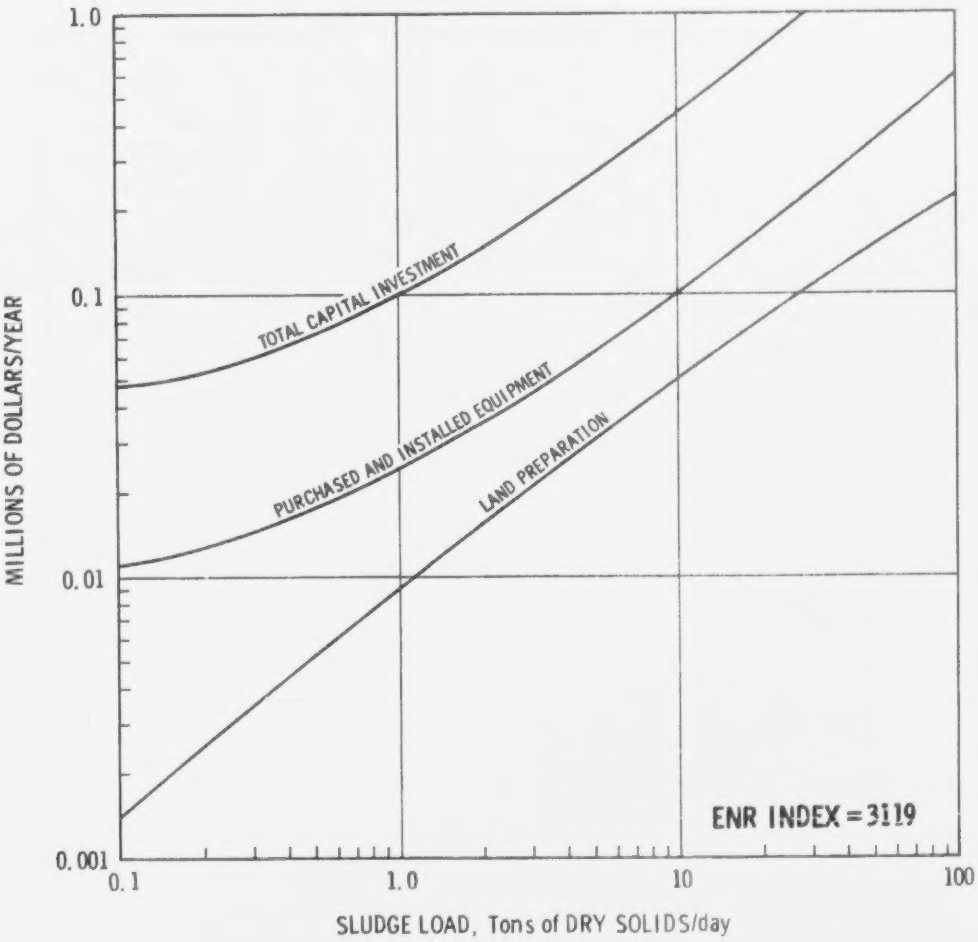
Item		Estimated cost[a]	
		Cost/yr	Cost ton dry solids
Maintenance and repair		$ 2,000	$0.60
Fuel cost, diesel @ 60¢/gal		7,200	2.16
gas @ 65¢/gal		11,700	3.50
Depreciation		5,000	1.49
Labor		13,800	4.14
Total	Diesel	29,000	8.39
	Gas	32,500	9.73

[a] Prices as of November 1977

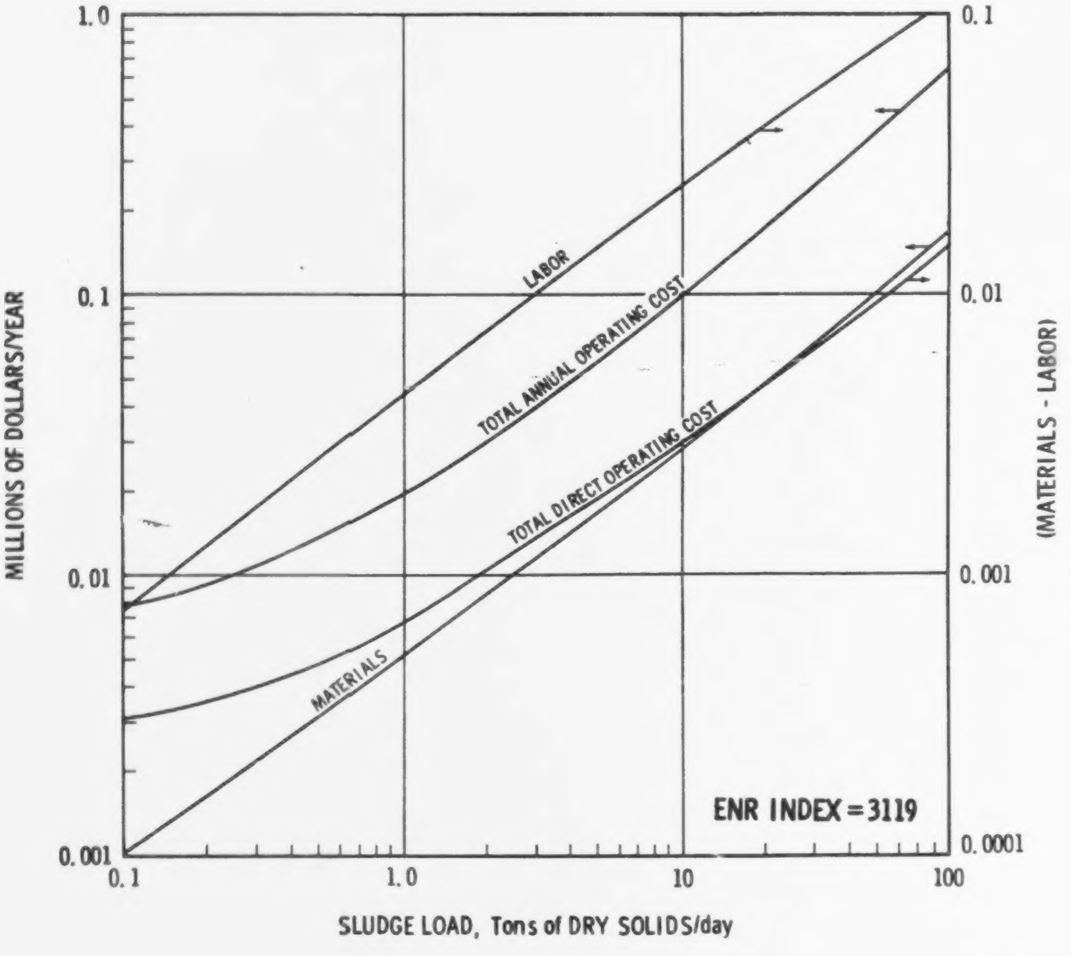
IV.7.5.4 References

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7.6 COMPOSTING

IV.7.6.1 Static Pile Composting Sludge [1]

Description. In static pile composting, sewage sludge is converted to compost in approximately eight weeks in a four-step process involving preparation, digestion, drying and screening, and curing.

In the preparation step, sludge is mixed with a bulking material such as wood chips or leaves to facilitate handling, provide the necessary structure and porosity for aeration, and lower the moisture content of the biomass to 60% or less. Following mixing, the aerated pile is constructed and positioned over porous pipe through which air is drawn. The pile is covered for insulation.

In the digestion step, the aerated pile undergoes decomposition by thermophilic organisms, whose activity generates a concomitant elevation in temperature to 60°C (140°F) or more. Aerobic composting conditions are maintained by drawing air through the pile at a predetermined rate. The effluent air stream is conducted into a small pile of screened, cured compost where odorous gases are effectively absorbed. After about 21 days, the composting rates and temperatures decline, and the pile is taken down, the plastic pipe is discarded, and the compost is either dried or cured depending upon weather conditions.

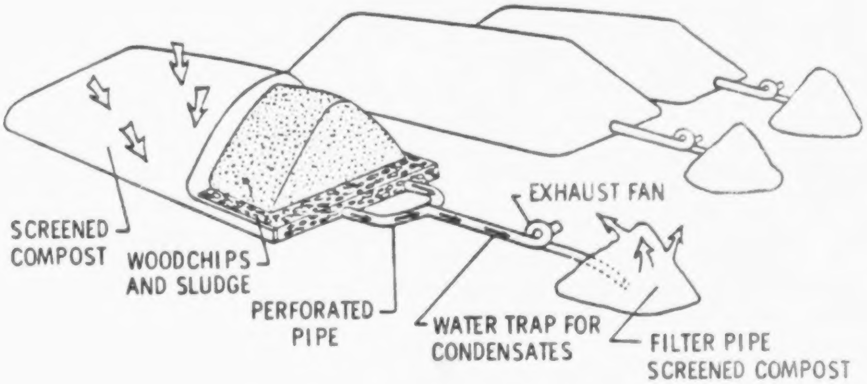
In the drying and screening step, drying to 40% to 45% moisture facilitates clean separation of compost from wood chips, which are recycled. The unscreened compost is spread out with a front-end loader to a depth of 12 inches. Periodically a tractor-drawn harrow is employed to facilitate drying. Screening is performed using a rotary screen.

In the curing step, the compost is stored in piles for about 30 days to assure that no offensive odors remain and to complete stabilization. The compost is then ready for utilization as a low-grade fertilizer, a soil amendment, or a material for land reclamation.

Two common modifications of static pile composting sludge are used: (1) extended high pile in which pile height is extended to 18 ft using a crane (still experimental), and (2) aerated extended pile in which each day's pile is constructed against the shoulder of the previous day's pile, forming a continuous or extended pile. These modifications result in savings of space and materials.

Equipment normally associated with static pile composting sludge includes the front-end loader or crane, 4-inch perforated plastic pipe, blower, timer, tractor (drawn), harrow, and rotary screen.

A flow diagram of the static pile composting of sludge is shown below.



Typical Design Criteria. Construction of the pile for a 10-dry-ton/d (43-wet-ton/d) operation involves the following steps: (1) a 6-inch layer of unscreened compost for base; (2) a 94-foot loop of 4-inch diameter perforated plastic pipe placed on top (0.25-inch hole diameter); (3) a 6-inch layer of unscreened compost or wood chips covering the pipe; (4) connection of the loop to a 1/3-hp blower using 15 feet of solid pipe fitted with a water trap to collect condensate; (5) a timer set for a cycle of 4 minutes on and 16 minutes off; (6) a blower connected to a covered scrubber pile (2 yd^3 wood chips covered with 10 yd^3 screened compost) using 16 feet of solid pipe; (7) placement of a wet sludge-wood chip mixture, in a volumetric ratio of 1:2.5, on a prepared base; and (8) placement of a 12-inch layer of screened compost on top for insulation. Air flow is 100 ft^3/h/ton of sludge. Approximately 3.5 acres of land area are required for processing 10 dry tons daily; the area includes a runoff collection pond; bituminous surface for roads, mixing, composting, drying, and storage; and an administration area. Pile dimensions are 53 ft x 12 ft x 8 ft high.

Costs. The following operating characteristics are assumed for the cost estimation presented in the tables on the following page and related to graphical form using scaling factors.

Operating characteristic	Assumed value
Quantity processed	10 dry tons/d (compost distribution will realize no net revenues or costs to the industrial)
Power for blower	1/3 hp
Front-end loader size	3.5-yd^3 bucket
Sewer line installation	400 ft of 8-in. sewer line @ $35/ft
Composting pad construction	grading, 12 in. crushed stone, 4 in. asphalt
Operation rate	8 hr/d, 7 d/wk
Staff	1 superintendent @ $7.50/hr, 4 equipment operators @ $6/hr
Labor cost	includes 5 wk off for paid sick leave, vacations, holidays; $400/person health insurance; 6% FICA; 0.3 man-yr of overtime
Equipment maintenance	6% purchase price
Estimated insurance	1% purchase price
Fuel cost, gasoline	57¢ gal, 1.1 gal/dry ton loading
diesel	41¢/gal, 3.5 gal/dry ton loading
electricity	2¢/kWh, 17.3 kWh/dry ton loading
woodchips	$3.50/$yd^3$

CONSTRUCTION COST

Item	Cost
Site development	
Asphalt pad (1.5 acre)	$ 83,800
Roads, administration	13,000
Electrical work	20,000
Sewer	14,000
Pond, drainage	28,000
Equipment	
Office trailer	5,000
Storage	1,500
Front-end loaders (2 pieces)	106,000
Screen	16,300
Tractor	4,700
Pickup	4,700
Blowers (33 pieces)	2,500
Construction cost	$299,500

OPERATION AND MAINTENANCE COST

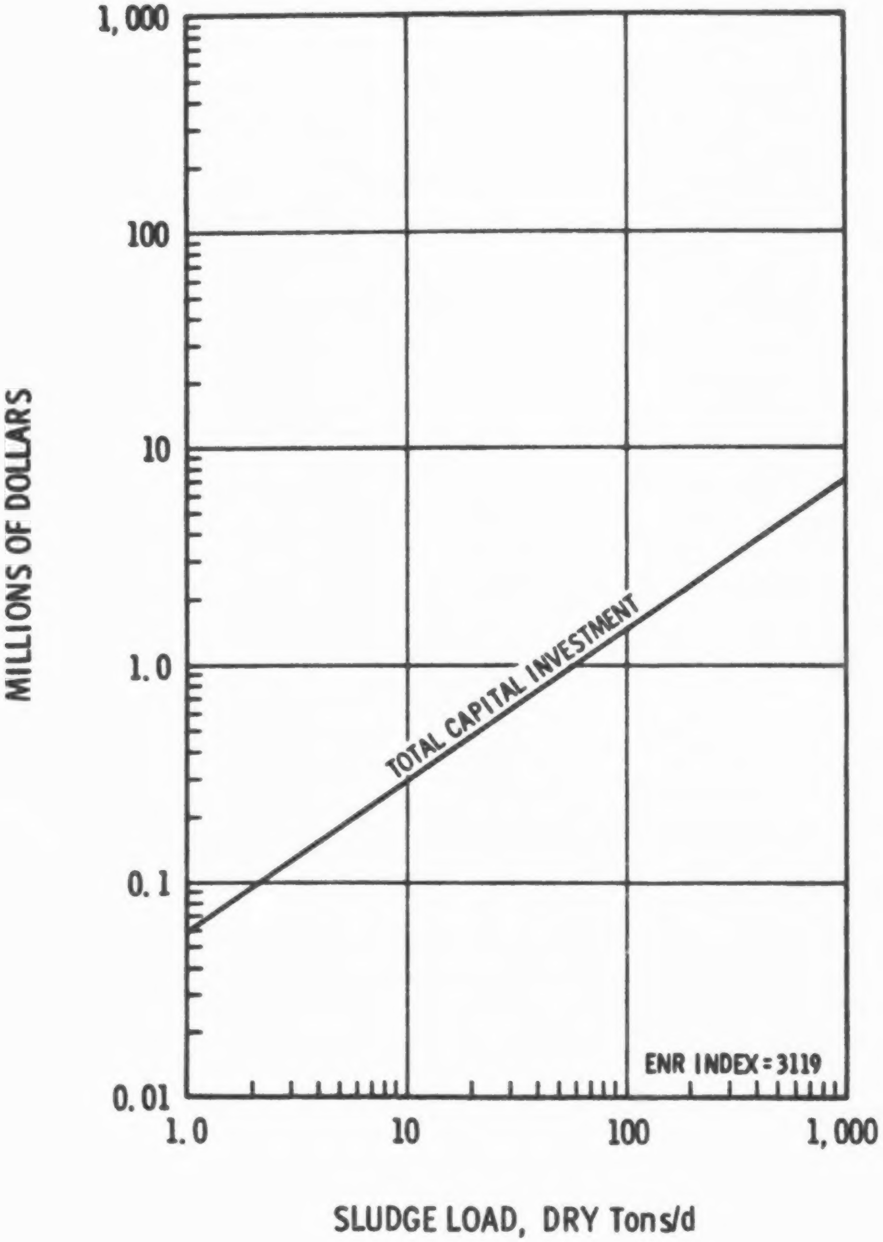
Item	Cost/yr	Cost/dry ton
Woodchips	$ 35,000	$ 9.60
Plastic pipe	12,200	3.34
Gasoline	2,300	0.63
Diesel	5,300	1.45
Electricity	1,500	0.41
Equipment maintenance	8,400	2.30
Equipment insurance	1,400	0.44
Pad, road maintenance	1,200	0.33
Water/sewer	500	0.14
Labor	77,500	21.23
Miscellaneous supplies	4,400	1.20
Total	$149,700	$41.01[a]

[a] O&M costs for a 50-dry-ton/d operation have been estimated to amount to $28/dry ton.

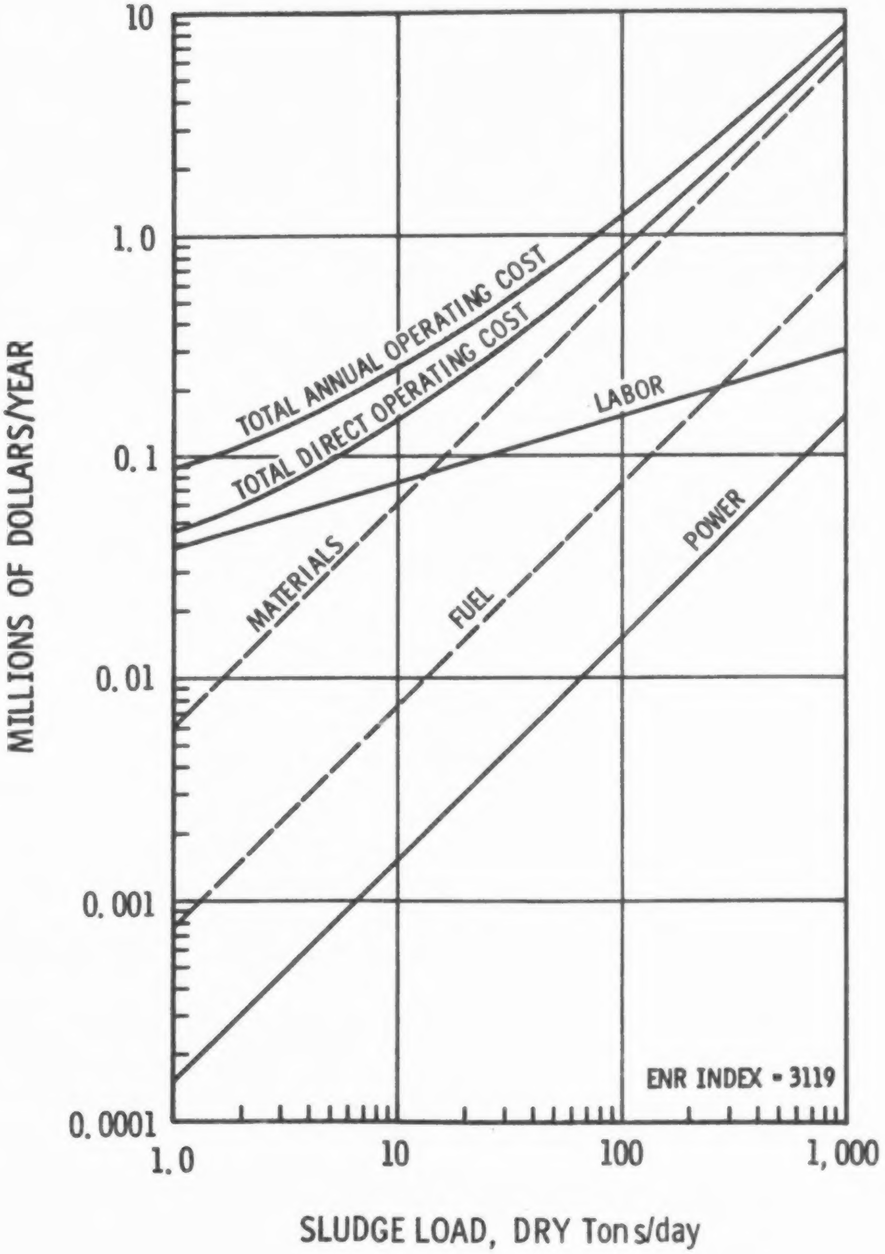
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7.6.2 Windrow Composting Sludge [1]

Description. In windrow composting, the piles are turned periodically to provide oxygen for the microorganisms to carry out the stabilization and carry off the excess heat generated by the process. When masses of solids are assembled, and conditions of moisture, aeration, and nutrition are favorable for microbial activity and growth, the temperature rises spontaneously. As a result of biological self-heating, composting masses easily reach 60°C (140°F) and commonly exceed 70°C (150°F). Peak composting temperatures approaching 90°C (194°F) have been recorded. Temperatures of 140°F to 160°F serve to kill pathogens, insect larvae, and weed seeds. Nuisances such as odors, insect breeding, and vermin harborage are controlled through rapid destruction of putrescible materials. Sequential steps involved in composting are preparation, composting, curing, and finishing.

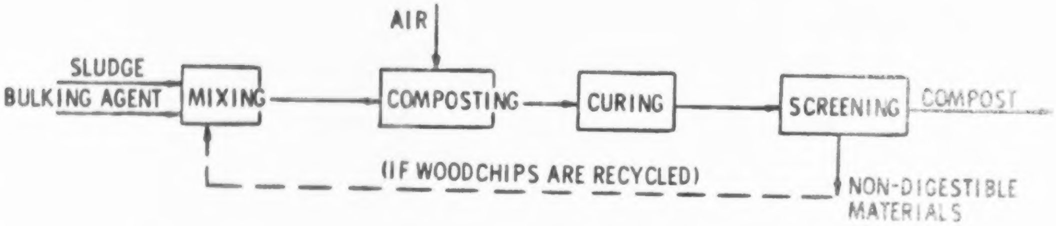
The preparation step involves adjusting the sludge properties in order to permit composting. To be compostable, a waste must have at least a minimally porous structure and a moisture content of 45% to 65%. Therefore, sewage sludge cake, which is usually about 20% solids, cannot be composted by itself but must be combined with a bulking agent, such as soil, sawdust, wood chips, refuse, or previously manufactured compost. Sludge and refuse make an ideal process combination. Refuse brings porosity to the mix, while sludge provides needed moisture and nitrogen; both are converted synergistically to an end product amenable to resource recovery. The sludge is suitably prepared and placed in piles or windrows.

In the composting step, the composting period is characterized by rapid decomposition. Air is supplied by periodic turnings. The reaction is exothermic, and wastes reach temperatures of 140°F to 160°F or higher. Pathogen kill and the inactivation of insect larvae and weed seeds are possible at these temperatures. The period of digestion is normally about six weeks.

Curing is characterized by a slowing of the decomposition rate. The temperature drops back to ambient, and the process is brought to completion. The period takes about two more windrow weeks.

In finishing, some sort of screening or other removal procedure is necessary if municipal solid waste fractions containing nondigestible debris have been included, or if the bulking agent such as wood chips is to be separated and recycled. The compost may be pulverized with a shredder, if desired.

Equipment normally associated with windrow composting sludge include front-end loaders, tractor (drawn), harrow and rotary screen. Equipment is currently being developed specifically for sludge composting. A flow diagram of a windrow composting sludge system is shown on the following page.



Typical Design Criteria. Typical design criteria for windrow composting of sludge are shown below. Sludge cannot be composted by itself but must be combined with a bulking agent to provide the biomass with the necessary porosity and moisture content.

Design criteria	Value
Land required	Approximately 1/3 acre/dry ton sludge daily production
Windrow size	4- to 8-ft high, 12- to 25-ft wide at base; length variable
Biomass specifications:	
Moisture content	45 - 65%
Carbon/nitrogen ratio	30 - 35%
Carbon/phosphorus ratio	75/1 - 150/1
Air flow	10 - 30 ft³/d/lb
Detention time	6 wk - 1 yr

Costs. Purchased equipment and installation cost for estimation of the total capital investment includes asphalt pads, roads, sewer, drainage pond, electrical work, and engineering. The following operating characteristics were assumed for cost estimation.

Operating characteristic	Assumed value
Service life	17 yr
Sludge production rate	900 lb/Mgal (dry solids) digested
Land requirement	0.35 acre/(ton/d)

Costs are based on composting of digested or raw biological sludge. Assumed land cost equals $10,500/acre.

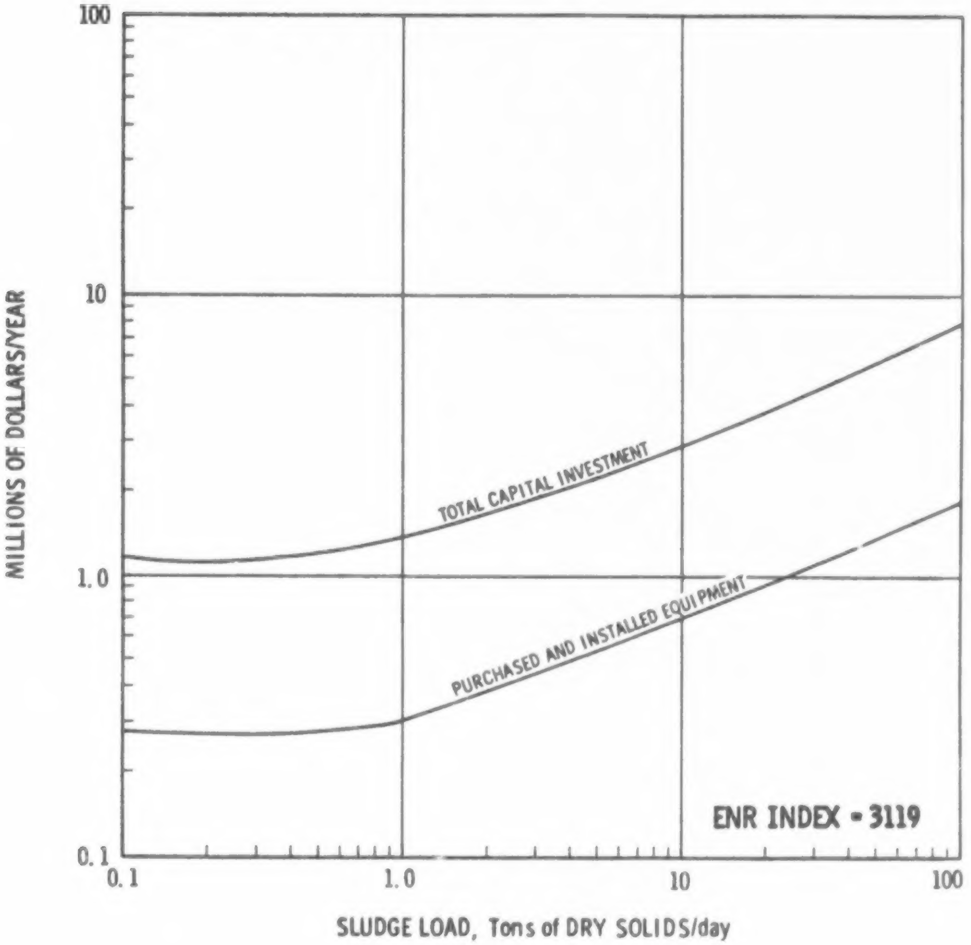
To adjust costs for composting rates different from 900 lb/Mgal, calculate the effective flow (Q_E) using the following equation:

$$Q_E = Q_{Design} \times \frac{\text{new design sludge mass}}{900 \text{ lb/Mgal}}$$

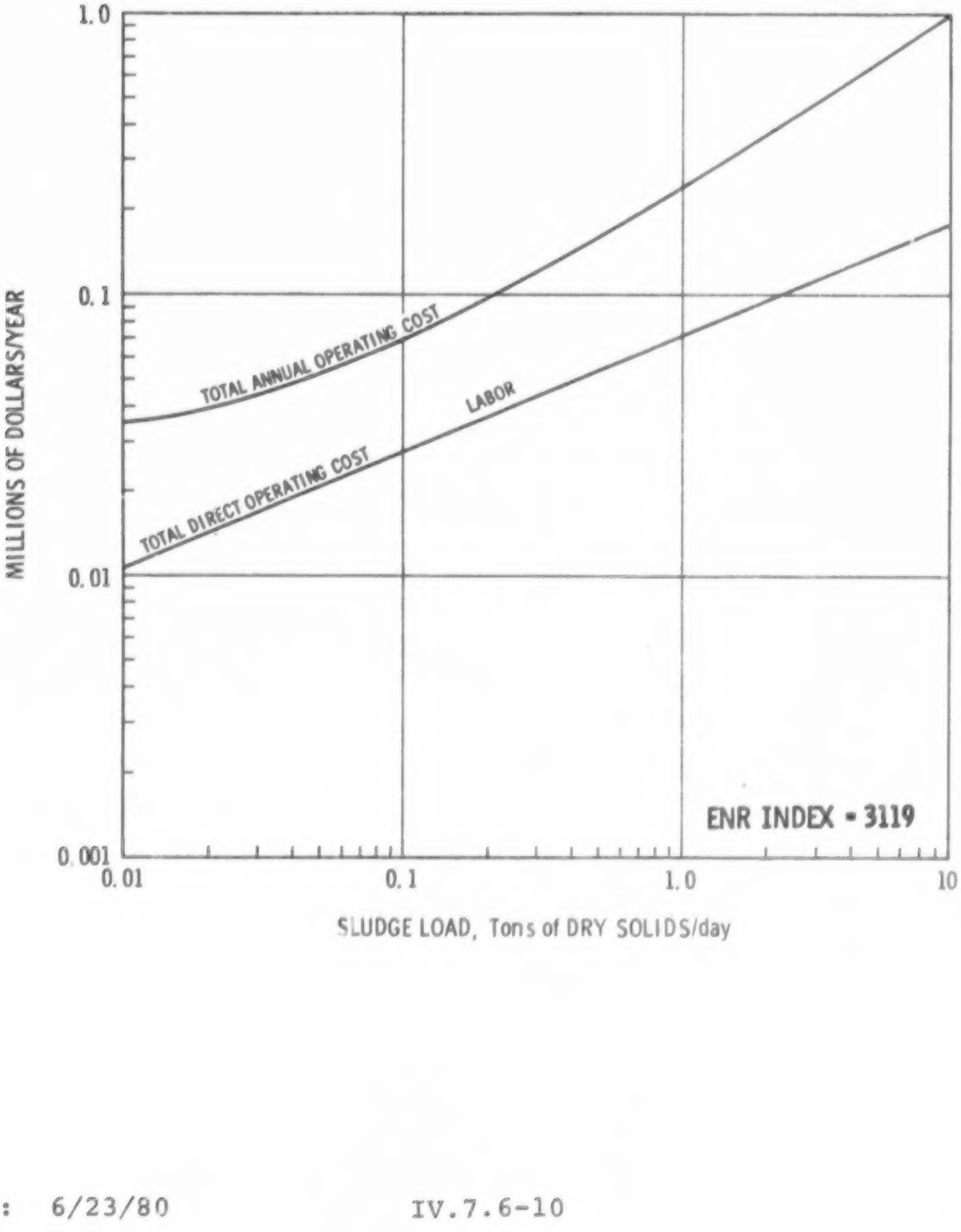
References.

1. Innovative and Alternative Technology Assessment Manual, EPA-430/9-78-009 (draft), U.S. Environmental Protection Agency, Cincinnati, Ohio, 1978. 252 pp.

TOTAL CAPITAL INVESTMENT



ANNUAL OPERATING COST



IV.7.7 DEEP-WELL INJECTION [1]

IV.7.7.1 Description

Deep-well disposal of industrial wastes, which has been practiced for over 25 years, is today a well-developed, carefully controlled, method of liquid waste disposal. All types of waste from a wide range of industries are being injected into deep wells. These wastes vary from relatively innocuous to highly toxic, from practically inert to highly reactive.

Basically, an industrial waste stream is deep-well injected only after all reasonable alternative disposal methods have been evaluated and found less desirable in terms of environmental protection and dependability. Chemicals found in waste streams that may have to be deep-well injected can be divided into organic and inorganic types. They may be liquids, gases, or solids; the gases and solids are either dissolved in the liquids or simply carried along by them.

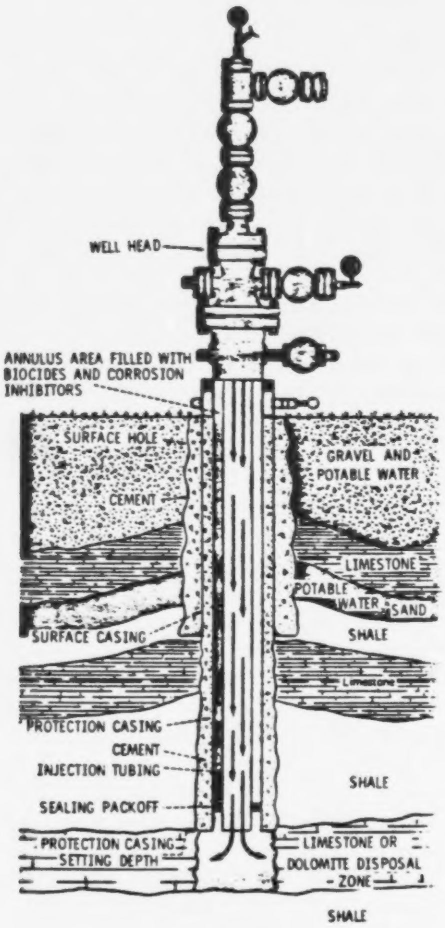
The organic chemicals include acids, such as maleic, formic, adipic, cresylic, salicylic, and acetic; alcohols, such as methanol, tertiary butanol, phenol, and isopropanol; solvents, such as acetone, toluene, xylene, formaldehyde, ethylbenzene, benzaldehyde, and methyl ethyl ketone; and other compounds, such as sodium naphthenate, sodium cresylate, calcium and sodium acetate, and large molecular structures such as styrene polymers and various polymeric resins.

Inorganic chemicals include acids, such as sulfuric, hydrochloric, and phosphoric; bases, such as sodium hydroxide; and salts, such as sodium chloride, sodium sulfate, sodium sulfide, sodium carbonate, sodium sulfite, arsenic sulfide, ammonium bisulfate, sodium bromide, and calcium carbonate.

A deep-well disposal system consists of a disposal zone, well, and surface facility for pretreating the waste liquids. The disposal zone must be located below potable-water aquifers, and isolated from them by thick, relatively impermeable and fracture-resistant strata such as shale, limestone, or dolomite. The zone must totally contain the waste liquids and have no other utility value. Deep-well disposal-zone depths vary from a few hundred to several thousand feet.

Disposal wells are constructed using oil- and gas-industry-proven technology, incorporating special adaptions for problems unique to waste injection. The key to a successful deep-well system is compatibility of the waste liquids with the materials of construction, the formation fluid, and the formation itself.

A diagramatic sketch of a deep-well disposal system is shown on the following page.


WELL HEAD
ANNULUS AREA FILLED WITH BIOCIDES AND CORROSION INHIBITORS
SURFACE HOLE
CEMENT
GRAVEL AND POTABLE WATER
LIMESTONE
POTABLE WATER
SAND
SURFACE CASING
SHALE
Limestone
PROTECTION CASING
CEMENT
INJECTION TUBING
SHALE
SEALING PACKOFF
PROTECTION CASING SETTING DEPTH
LIMESTONE OR DOLOMITE DISPOSAL ZONE
SHALE

IV.7.7.2 Typical Design Criteria

The design of a given deep-well disposal system must be based upon the type of disposal zone, its permeability and storage capacity, the volume of waste to be injected, the rate at which it is to be injected, and the type of waste.

IV.7.7.3 Costs

Several factors determine the economics of a deep-well disposal system. The major factors are depth of disposal zone, proposed rate of injection, preinjection treatment required, formation injectivity, power costs, labor costs, and contract drilling costs.

Costs vary with each operation, but, in general, installation of a deep-well disposal system costs about $750,000 to $1,000,000 in 1979 dollars.

For a well that is one mile deep and has an injection rate of 6 Mgal/mo (140 gpm), maintenance costs in 1979 dollars are about $60,000/yr/active well; energy costs run about 2¢/gal or less, depending on utility rates. If neutralization of the waste is required prior to injectuon, this could cost as much as $150,000/yr/active well.

IV.7.7.4 References

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APPENDIX A

ECONOMIC ASSUMPTIONS

COST COMPONENTS

Total capital investment and annual operating cost curves were generated for wastewater treatment technologies grouped into the following classifications: wastewater conditioning, primary wastewater treatment, secondary wastewater treatment, tertiary wastewater treatment, sludge treatment, and disposal. The following sections discuss general factors contributing to these costs and assumptions used in this manual to generate total capital investment and annual operating cost curves, respectively.

Total Capital Investment

Total capital investment, the total amount of money to supply the necessary treatment facilities plus that required for initial operation of the facilities, is herein the sum of the fixed capital investment and a small amount of working capital. Fixed capital investment was divided into direct cost components and indirect cost components as shown in Table A-1, which lists the major components of the total capital investment and the assumptions used in estimating the total capital investment from purchased and installed equipment values presented in the literature. Assumed values are similar to those presented in Reference 1 with adjustments resulting in a lower percentage for working capital.

Total Annual Operating Cost

The total annual operating cost curve is presented on the same graph as cost curves for labor, materials, chemicals, power, and fuel, when applicable, and a curve for total direct operating cost (calculated as the sum of the costs for labor, materials, chemicals, power, and fuel). Table A-2 lists the major components of the annual operating cost and the assumptions used in estimating the annual operating cost from labor, materials, chemical, power, and fuel values presented in the literature. Cost components marked with an asterisk are those which have been plotted, where appropriate.

TABLE A-1. COST ASSUMPTIONS AND BREAKDOWN OF TOTAL CAPITAL INVESTMENT

Cost component	Assumed value
Fixed capital investment:	
Direct costs	
• Purchased equipment and installation	PE&I
• Instrumentation and controls	10% PE&I
• Piping	21% PE&I
• Electrical equipment and materials	13% PE&I
• Buildings	26% PE&I
• Yard improvements	7% PE&I
• Service facilities	41% PE&I
Indirect costs	
• Engineering and supervision	29% PE&I
• Construction expenses	32% PE&I
• Contractor's fees	7% PE&I
• Contingency	27% PE&I
Total fixed capital investment	3.13 PE&I
Working capital	0.47 PE&I
Total capital investment	3.6 PE&I

TABLE A-2. COST ASSUMPTIONS AND BREAKDOWN OF TOTAL ANNUAL OPERATING COST

• Labor	(L)
• Materials	(M)
• Chemicals	(C)
• Power	(P)
• Fuel	(F)
• Total Direct Operating cost (TDC) = (L + M + C + P + F)	
Plant overhead	- (60% L)
Taxes and insurance	- (2% FCI)
General and administrative expenses	- (40% L)
Depreciation	- (10% FCI)
Interest on working capital	- (12% WC)
Annual operating cost	- (TDC + L + 12% FCI + 12% WC)

The following paragraphs briefly describe the individual cost components [2].

Labor costs. Labor costs include the manpower required to operate and maintain the facility, plus such support tasks as supervision and administration, clerical work, laboratory work, and yard work.

Material costs. Material costs include the various materials required for routine maintenance of the facility. Examples are paint, grease, and replacement parts. Such costs for mechanical equipment, such as pumps and aerators, are generally higher than those for structural equipment.

Chemical costs. Various unit processes require chemicals. The quantities of chemicals required are based on the assumed loading rates. The chemical costs are generated from the unit price for each chemical and the assumed loading rate.

Power costs. Each mechanical operation at a treatment facility, such as pumping, mixing, and aeration, consumes energy. Horsepower requirements are converted into electrical units (kWh) and a corresponding cost is developed using a unit price per kWh.

Fuel costs. Energy other than electricity is required for some processes (such as incineration of sludge). In these cases, the amount of energy (fuel) required is computed and the cost determined on the basis of an assumed cost per unit.

DEVELOPMENT OF COST CURVES

A large number of both construction and operating and maintenance cost curves were used as a basis for development of total capital cost curves and annual operating cost curves for this manual. These construction (purchased equipment and installation) cost curves and operating and maintenance cost curves were derived from the Innovative and Alternative Technology Assessment Manual [3] and the Areawide Assessment Procedures Manual [2], and were originally developed primarily from 1976 cost data and municipal wastewater treatment technology. Cost data indexed to years other than 1976 were also used as a basis for development of cost curves in this manual. The costs from various years were indexed to September 1979 using both Engineering News Record (ENR) cost indices (ENR index equals 3119) and direct ratios. The ENR construction cost index was used to index purchased and installed equipment to 1979 and thus to determine total capital investment cost. Additionally, ENR common labor and material cost indices were used to scale up labor and material costs; direct ratios were used to scale up power and fuel costs; and chemical costs were assumed to be unchanged.

As previously mentioned, the base for costs presented in this manual is generally 1976 and these costs were subsequently indexed to 1979. From Reference 3, the original base is primarily an ENR index of 2475 and those assumptions listed in Table A-3.

TABLE A-3. COST AND DESIGN BASES FOR REFERENCE 3 1976 COST DATA

Cost basis	Assumed value
ENR index	2475, September 1976
Labor rate[a]	$7.50/hr including fringe benefits
Energy costs	
Electric power	$0.02/kWh
Fuel oil	$0.37/gal
Gasoline	$0.60/gal
Land	$1,000/acre
Chemical costs	
Liquid oxygen	$65/ton
Methanol	$0.50/gal
Chlorine, 150-lb cylinders	$360/ton
1-ton cylinder	$260/ton
tank car	$160/ton
Quicklime	$25/ton
Hydrated lime	$30/ton (as CaO)
Polymer (dry)	$1.50/lb
Ferric chloride	$100/ton
Alum	$72/ton
Activated carbon (granulated)	$0.50/lb
Sulfuric acid (66° Be)	$50/ton
Sodium hexametaphosphate	$0.25/lb
SO_2, 150-lb cylinders	$450/ton
1-ton cylinder	$215/ton
tank car	$155/ton

Design basis

Construction costs and operation and maintenance costs are based on design average flow unless otherwise noted.

Operation and maintenance costs include:

Labor costs for operation, preventative maintenance, and minor repairs.
Materials costs to include replacement parts and major repair work (normally performed by outside contractors).
Chemical costs.
Fuel costs.
Electrical power costs.

Construction costs do not include external piping, electrical, instrumentation, site work, contingency, or engineering and fiscal fees.

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[a] Based on 1,500 hr/man-year and represents: a 5-day work week; an average of 29 days for holidays, vacation, and sick leave; and 6.5 hours of productive work time per day.

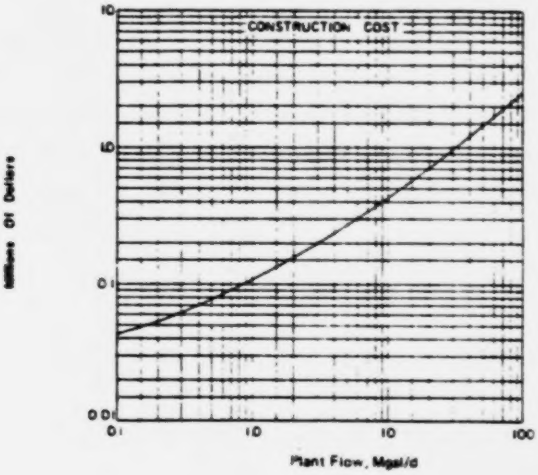
As an illustration of the procedure used in indexing and scaling up the 1976 cost curves, the following example is presented. Reference 3 gave this information for primary circular clarification (with pump).

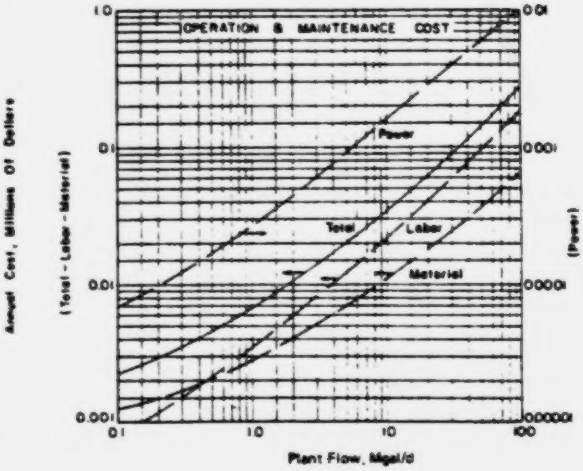
COSTS - Service Life = 50 years; ENR = 2475.

Design Basis:

1. Clarifier designed for surface overflow rate of 800 gal/d/ft² (based on average Q).
2. Costs include primary sludge pumps; sludge concentration of 4 percent solids; pump head assumed as 10 ft TDH.
3. Power Cost = $0.02/kWh
4. To adjust costs for alternative surface overflow rate, enter flow at effective flow (Q_E)

$$Q_E = Q_{Design} \times \frac{800 \text{ gal/d/ft}^2}{\text{New Design Surface Overflow Rate}}$$





The total capital investment and purchased and installed equipment costs in 1979 dollars were estimated from the information above. Using the ratio of 1979 to 1976 ENR indices (3119/2475 = 1.26), the purchased and installed equipment cost was plotted in 1979 dollars. From the purchased and installed equipment cost, total capital investment was then estimated by applying the factor of 3.6 shown in Table A-1.

Development of total direct operating and annual operating cost curves required the use of individual indices for power, labor, and material shown below. The following example shows that costs were read from the graph for each of these components at flowrates corresponding to 0.1, 1.0, 10, and 100 Mgal/d. Each cost component was indexed to 1979 by either an ENR index or direct ratio.

These components were summed to give the total direct operating cost in 1979 dollars, the factors presented in Table A-2 were used to derive total annual operating cost, and the data were used to plot curves for power, labor, materials, total direct operating cost, and total annual operating cost in estimated 1979 dollars.

EXAMPLE TOTAL ANNUAL OPERATING COST CALCULATION
FOR PRIMARY CIRCULAR CLARIFICATION WITH PUMP

		Flowrate, Mgal/d			
Cost component	Factor	0.1	1.0	10	100
Labor	(L, from graph)	-	0.003	0.02	0.18
	(L x 1.21)[a]	-	0.0036	0.024	0.22
Materials	(M, from graph)	0.0013	0.0028	0.01	0.068
	(M x 1.39)[b]	0.0018	0.0039	0.014	0.095
Chemicals	(C, from graph)				
Power	(P, from graph)	0.0007	0.00026	0.0015	0.01
	(P x 1.75)[c]	0.00012	0.00046	0.0026	0.0175
Fuel	(F, from graph)				
	(F x 2.16)[d]				
Total direct operating cost	(L+M+C+P+F)	0.002	0.008	0.041	0.33
Total annual operating cost	(TDC[e]+L+0.12 TCI[f])	0.026	0.066	0.281	1.97

Note: Blanks indicate not applicable.

[a] Based on ENR common labor index. The ratio of September 1979 to 1976 indices is $\frac{5844}{4815}$ or 1.21.

[b] Based on ENR materials index. The ratio of September 1979 to September 1976 indices is $\frac{1449}{1040}$ or 1.39.

[c] Based on a direct ratio of 1979 costs to 1976 costs equal to approximately $\frac{\$0.035/\text{kWh}}{\$0.02/\text{kWh}}$ or 1.75.

[d] Based on a direct ratio of 1979 to 1976 costs equal to approximately $\frac{\$0.80/\text{gal fuel oil}}{\$0.37/\text{gal fuel oil}}$ or 2.16.

[e] TDC (total direct operating cost).

[f] Total capital investment.

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